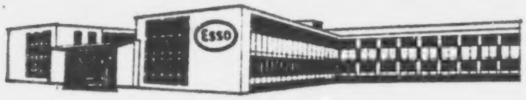
EPA R 2-72.020



ESSO RESEARCH CENTRE

ABINGDON BERKSHIRE

PB 211 438

FINAL REPORT

OAP CONTRACT CPA 70-46

JUNE 22 1970 TO MARCH 1972

WORK PERFORMED FOR
ENVIRONMENTAL PROTECTION AGENCY
CONTROL SYSTEMS DIVISION

STUDY OF CHEMICALLY ACTIVE FLUID BED GASIFIER FOR REDUCTION OF SULPHUR OXIDE EMISSIONS

Report By: J.W.T.Craig
G.L.Johnes
G.Moss
J.H.Taylor
D.E.Tisdall

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U S Department of Commerce
Springfield VA 22151

June 1972

347

BIBLIOGRAPHIC DATA SHEET

1. Report No.
EPA-R2-72-020

2.

3. Recipient's Accession No.
PB-211 438

4. Title and Subtitle
Study of Chemically Active Fluid Bed Gasifier for Reduction of Sulphur Oxide Emissions

5. Report Date
June 1972

6.

7. Author(s)
J.W.T. Craig, G.L. Johnes, G. Moss, et al

8. Performing Organization Rept. No.

9. Performing Organization Name and Address

ESSO RESEARCH CENTRE
ABINGDON BERKSHIRE
ENGLAND

10. Project/Task/Work Unit No.

11. Contract/Grant No.
CPA 70-46

12. Sponsoring Organization Name and Address

ENVIRONMENTAL PROTECTION AGENCY, OAP
Control Systems Division
Research Triangle Park, N. C. 27711

13. Type of Report & Period Covered
Final

14.

15. Supplementary Notes

16. Abstracts

The first phase is presented of an overall project which has the objective of demonstrating and evaluating the Chemically Active Fluid Bed (CAFB) process in a converted power generation boiler of approximately 150 megawatt capacity. The current work objectives of this phase I study were as follows: a) survey operating characteristics of two limestones and one residual fuel oil to determine their suitability for the CAFB process; b) select the best of the two stones for further testing; c) construct two new batch reactor CAFB laboratory experimental units; d) measure the effects of important operating variables on the CAFB process using batch reactor experiments with the selected oil-limestone combination; e) operate for 200 hours of gasification a continuous CAFB gasification pilot plant designed, constructed, and paid for by Esso. These experiments in batch reactors and operation of a continuous, two-reactor, pilot plant have greatly increased the understanding of the (CAFB) process for heavy fuel oil gasification and desulphurisation in a fluidized bed of lime particles. The experiments have demonstrated a range of conditions over which sulphur removal of 95% or more can be achieved while producing a hot fuel gas which burns readily in a burner of simple design.

17. Key Words and Document Analysis. 17a. Descriptors

Air pollution
Sulfur oxides
Fuel oils
Combustion
Fluidized bed processors
Chemical reactors
Pilot plants
Gasification
Desulfurization
Regeneration (engineering)
Limestone
Vanadium

17b. Identifiers/Open-Ended Terms

Air pollution control
Chemically Active Fluid Bed (CAFB)
Air pollution control

17c. COSATI Field/Group 13B, 7D

18. Availability Statement
Unlimited

19. Security Class (This Report)
UNCLASSIFIED

20. Security Class (This Page)
UNCLASSIFIED

21. No. of Pages
349

22. Price
$6.50

FORM NTIS-35 (REV. 3-72)

USCOMM-DC 14952-P72

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I Summary

Experiments in batch reactors and operation of a continuous, two-reactor, pilot plant have greatly increased our understanding of the Chemically Active Fluid Bed (CAFB) process for heavy fuel oil gasification and desulphurisation in a fluidised bed of lime particles. These experiments have demonstrated a range of conditions over which sulphur removal of 95% or more can be achieved while producing a hot fuel gas which burns readily in a burner of simple design.

A. Fuel and Limestone Screening

Preliminary screening tests in batch reactors built prior to this contract established that a Venezuelan fuel oil available in the United States could be simultaneously gasified and desulphurised satisfactorily by the CAFB technique. A comparison of two limestones suggested by NAPCA (now OAP) revealed that one, BCR 1691, was satisfactory for CAFB operation, though it would require higher replacement rates for equal sulphur removal because it contained less calcium oxide than the previous standard, a U.K. Denbighshire stone. Preliminary indications were that the stones were equivalent at equal CaO replacement rates. A second stone, BCR 1690, was inferior as a sulphur removal agent even at equal CaO replacement rates. It also was subject to greater attrition rates and showed a tendency to agglomerate. No further work was performed on BCR 1690.

B. Batch Reactor Variable Study

A batch reactor variable study in two new experimental units screened the effects of major process variables with the Venezuelan fuel oil (2.3% S, 350 ppm Vanadium) and the U.S. limestone BCR 1691. The effects on sulphur removal of limestone particle size, bed depth, superficial gas velocity, air/fuel ratio, and temperature were explored in tests with fresh lime beds. Bed depth, lime replacement rate, the amount of sulphur exposure per gasification cycle, and particle size were further studied in a series of cyclic tests using alternate cycles of gasification and regeneration.

Gasification temperature and air fuel ratio were found to interact strongly. Conditions that produced an adequate rate of carbon burning near the air distributor were required to avoid rapid loss of sulphur removal activity by excessive carbon laydown. With gasification temperatures of 870°C or higher, air/fuel ratios as low as 20% of stoichiometric could be used with good (>95%) sulphur removal efficiency. Temperatures down to 800°C could be used provided air/fuel ratio were high enough i.e. above 27% of stoichiometric. On the other hand, sulphur removal efficiency declined at temperatures above 900°C.

Sulphur removal activity of fresh beds correlated with the number average particle size of the lime.

It appears that, at least at low lime utilisations, sulphur removal efficiency varies inversely with the square of particle size. However, in cyclic testing or continuous operation, consideration must be taken of the effect of fines losses on performance. A limestone with a large fraction of less than 600 micron particles undergoes greater losses during calcination in the gasifier and gives poorer performance than a stone with larger initial particle size but with fewer losses. Extensive testing was done with a 300 - 3175 micron size range stone but a size range of 600 - 3175 microns is preferred.

The effects of bed depth and superficial gas velocity on sulphur removal efficiency can be correlated by their ratio, the superficial gas residence time above the fuel injector. As a first approximation, a first order reaction model expressed the variation of sulphur removal with gas residence time in a fresh bed. However, in cyclic tests, the deactivation of the lime with sulphur exposure must also be considered and gas residence time plays a less significant part. The empirical equation which fits the batch unit cyclic test data is:

$$\text{Sulphur Removal Efficiency, \%,} = 100 \left[1 - \frac{1}{\text{EXP}\left(\frac{9.3\, t^{.6} m^{.95}}{d^{1.9}}\right)} \right]$$

where t = gas residence time, sec above fuel injector

m = Lime replacement rate, wt CaO/wt S

d = % Sulphur differential, wt S passed into 100 lb lime.

This model implies a fluid bed depth of about 3.4 ft for 95% sulphur recovery with stoichiometric replacement of BCR 1691 stone at 6 ft/sec velocity.

Vanadium is removed from the fuel oil by the lime during gasification. In some cyclic test series vanadium removal from the fuel was essentially complete. In other cases less than 100% removal was experienced. Effectiveness of vanadium removal correlates with regeneration temperature, which implies that a vanadium fixation step is needed in addition to the preliminary vanadium absorption.

Lime losses from the bed in batch unit studies are sensitive to bed depth. They vary from about 1 to 9 lb/hr - ft^2 at 6 ft/sec superficial gas velocity. Similar loss rates have been observed in the continuous pilot plant. In the pilot plant maintenance of cyclone efficiency is a major factor in keeping losses low.

Empirical equations were obtained to describe the variation in composition of the gasifier product gas with temperature and air/fuel ratio. The CO/CO_2 ratio increases with temperature as does the fraction of hydrogen and hydrocarbons heavier than C_4. Increasing air fuel ratio increases the preferential oxidation of the fraction heavier than C_4

C. Continuous Pilot Plant Operation

Three runs of the continuous gasifier pilot plant have been completed. The first run demonstrated basic operability of the process and showed that continuous gasification, regeneration, solids circulation, and sulphur removal was feasible. An average of 90% sulphur removal from the fuel gas was maintained for a continuous 87 hour period while producing a regenerated gas containing 8 to 10% SO_2. Boiler operation showed that the gasifier produced a hot, combustible fuel which ignited readily and burned with a clean luminous flame.

During this run it was observed that carbonaceous deposits accumulated in the cyclones and gasifier outlet ducts during the run, and this accumulation caused a gradual increase of pressure drop between gasifier and boiler. The deposit was removed by air burnout, but excessive temperature during this burnout damaged the initial cyclone dip tubes made of stainless steel.

The second run was an attempt to operate the gasifier without internal dip tubes in the gasifier outlet cyclones.

A continuous gasification period of 91 hours was maintained with nearly 100% removal of sulphur from the fuel gas. However, absence of the cyclone tubes, coupled with high lime attrition rates caused by modified high velocity fuel injectors, led to high lime losses and required unreasonably high replacement rates.

The third run employed new cyclone dip tubes which could be steam cooled during decoking operations. A run with over 200 hours of gasification was carried out, using the steam cooling together with inert gas recycle to control temperature when decoking.

This third run demonstrated the following features:-

- Continuous gasification periods up to 64 hours without decoking.
- Essentially 100% sulphur removal at 100 to 200% of stoichiometric lime replacement rate and total bed depth of 19 inches.
- Greater than 95% sulphur removal at 70% of stoichiometric lime replacement and 22 inches total bed depth.
- Maintenance of regenerator gas concentrations of 6 to 9% SO_2.
- Automatic control of regenerator temperature by solids circulation rate.

- Regulation of gasifier temperature by flue gas recycle rate.
- Trouble-free operation of the solids circulation system.

These pilot plant runs pointed out several areas which require additional study to improve long term process operability.

- Decrease rate of solids deposition in cyclones and transfer ducts.
- Reduce lime attrition and solids carry-over to boiler.
- Find improved material of construction for cyclone internals to reduce time required for decoking by allowing higher burn-out temperature.

II Body of Report

A. Introduction and Objectives

Exploratory work at the Esso Research Centre, between 1966 and 1970, had shown a range of conditions under which sulphur is absorbed in a fluidised bed of lime in which fuel oil is undergoing partial combustion and gasification. The possibility of regenerating the lime also had been demonstrated. Thus the framework had been established for a process to eliminate SO_2 pollution during combustion of high sulphur fuel oils. The process is designated CAFB for Chemically Active Fluid Bed.

The process employs a shallow (about 2 feet deep) fluidised bed of lime particles in a gasifier to partially oxidise, to gasify, and to desulphurise the heavy fuel oil. The hot gasified, sulphur-free fuel passes from the fluid bed to equipment in which it is fully combusted, such as the burners of a boiler. Calcium sulphide, which forms in the gasifier fluid bed, is converted back to calcium oxide in a separate fluid bed regenerator by contact with air. Hot solids circulate continuously between gasifier and regenerator fluid beds. A purge stream of lime removed from the regenerator is replaced by an equivalent quantity of limestone to maintain activity of the bed for sulphur absorption.

The sulphur leaves the system as a concentrated (about 10%) SO_2 stream from the regenerator. This sulphur can be recovered in several ways; for example, by conversion to sulphuric acid or by direct reduction to elemental sulphur.

In addition to eliminating SO_2 pollution, the process has the additional benefit of removing vanadium before the final combustion stage, and thus reducing or eliminating this source of high temperature corrosion in boiler superheaters.

In June 1970, Esso Research Centre began work on a contract with the United States National Air Pollution Control Administration (now Office of Air Programs) to study the CAFB process for possible applications in the United States. This is the final report on work done under that contract in the period from June 22 1970 through March 1972.

1. Background Information

The original CAFB work at Esso Research Centre was directed at fully combusting oil fired fluid beds with the objective of achieving sulphur removal along with the high heat transfer rates of fluid bed boilers. It soon became evident that by using lime as the fluid bed medium, benefits in pollution abatement could be obtained which potentially were more important than the improved heat transfer.

It was found that sulphur was trapped by lime both under fully combusting conditions and under gasifying conditions. In the latter case the oil is contacted with substoichiometric amounts of air, and the gasifier product is a hot, sulphur-free fuel gas which can be burned in a conventional boiler.

Under fully combusting conditions the sulphur reacts with lime to form calcium sulphate.

$$H_2S + 2O_2 + CaO \longrightarrow CaSO_4 + H_2O$$

The sulphate can be reduced to SO_2 and lime with fuel oil. With hydrogen the reduction reaction is :

$$CaSO_4 + H_2 \longrightarrow CaO + SO_2 + H_2O$$

When fuel is combusted under substoichiometric conditions, the sulphur is trapped by lime as calcium sulphide:

$$H_2S + CaO \longrightarrow CaS + H_2O$$

The lime can be regenerated by oxidising the CaS with air:

$$CaS + \frac{3}{2} O_2 \longrightarrow CaO + SO_2$$

Each pair of reactions forms the basis of a regenerative lime process for avoiding SO_2 pollution when using high sulphur fuel oil. The gasifier process has the added advantage that it is potentially applicable to existing power plant installations.

The early work was described in a technical proposal to NAPCA (2) in 1969. Later work obtained in the period between submission of the proposal and beginning of work under contract CPA-70-46 is summarised in Appendix B of this report. A more complete discussion of CAFB chemistry appears in Appendix A.

Although the CAFB process uses continuous circulation of solids between two fluid bed reactors, the experimental work used a single reactor which was cycled between gasification and regeneration conditions. The earlier tests used reactors without cyclones and were restricted to low gas velocities to prevent solids carryover. Later versions of the apparatus employed internal cyclones and were able to use higher gas velocities. Temperatures in the range of 800-900°C were found best for sulphur removal during gasification. The regeneration of sulphided lime occurred between 1,000 and 1,100°C.

With the low gas velocity reactor it was necessary to underfire - that is, to burn some fuel in the plenum space beneath the air distributor. If underfiring were not used, the velocity of the cold entering air was insufficient to fluidise the solids in the vicinity of the distributor, and poor desulphurising efficiency was obtained. Even with underfiring, desulphurising efficiency declined severely as air to fuel ratio decreased below 30% of stoichiometric. When reactors with cyclones were employed, higher total gas rates could be employed, and underfiring was no longer required. Under these conditions, good sulphur removal was obtained at air rates down to nearly 20% of stoichiometric.

When a single bed of lime is cycled between gasification and regeneration conditions there is a gradual loss of sulphur removal efficiency. If a portion of the lime is replaced with fresh limestone at the beginning of each cycle, the efficiency will line out at some level which depends on the quantity of lime replaced. A curve of lined out efficiency vs. lime replacement rate was established which indicated that replacement of about one mole of CaO per mole of sulphur fed would maintain the sulphur removal efficiency at close to 100%. The bed life tests on which this replacement study was based were made with underfiring and about 30 to 35% of stoichiometric air.

These life tests also demonstrated that the lime retained much of the vanadium and some of the sodium from the fuel oil. About 20% of the sodium was picked up. Vanadium retention was variable and amounted to 100% in some instances. A fuller presentation of results appears in Appendix B.

2. Overall Project Objectives

Work under this contract constitutes Phase I of an overall project which has the objective of demonstrating and evaluating the CAFB process in a converted power generation boiler of approximately 150 megawatt capacity, situated in the United States. The overall project is conceived as being a six phase operation, with the following broad objectives for each phase:

Phase I
- Laboratory Tests in batch CAFB reactors to select suitable fuel and limestone combinations and to screen critical process variables.
- Construction and commissioning of a continuous pilot plant of about 7 million Btu/Hr capacity.

Phase II
- Operation of continuous pilot plant to provide process data and knowhow for design of conversion of commercial unit.
- Development of conceptual design for conversion and economic evaluation of process viability.

Phase III
- Full study design of proposed conversion.
- Additional laboratory and pilot plant studies to answer specific design questions.

Phase IV
- Detail design of conversion.

Phase V
- Boiler conversion and commissioning

Phase VI
- Boiler operation in normal duty while evaluating long term performance of CAFB gasifier.

The Phase I experimental programme is now complete.

3. Current Work Objectives

The specific objectives of this Phase I study have been as follows:

a. Survey operating characteristics of two limestones and one residual fuel oil selected by NAPCA (now OAP) to determine their suitability for the CAFB process.

b. Select the best of the two stones for further testing.

c. Construct two new batch reactor CAFB laboratory experimental units.

d. Measure the effects of important operating variables on the CAFB process using batch reactor experiments with the selected oil-limestone combination.

e. Operate for 200 hours of gasification a continuous CAFB gasification pilot plant designed, constructed, and paid for by Esso.

B. Apparatus, Materials and Procedures

Apparatus and procedures used in this study are described in this section. The apparatus includes the Esso batch unit used for the early limestone and fuel oil screening. The two new batch reactors used for variable studies, a cold solids model of the continuous gasifier, and the continuous gasifier pilot plant itself.

Materials include the various limestone and fuel oils tested, and a brick dust used to simulate hot lime in the cold gasifier model.

1. Batch Reactors

Two batch reactors built by Esso were in existence when work on this contract began. These units were used in the initial limestone and fuel oil screening work while two new batch units were being designed and constructed for the variable studies.

a. The 1-A Batch Reactor

The batch reactor used in the CAFB limestone and oil screening tests is designated No. 1-A. The lower section of this unit was the first CAFB reactor used extensively at the Esso Research Centre. In those original tests, however, it was not fitted with cyclones, and therefore was used only in low gas velocity experiments. Before the current work began the unit was fitted with an expanded upper section and an internal cyclone. This cyclone, which is closed at the bottom, traps fines but does not return them to the fluid bed. The quantity of fines is withdrawn and measured at the end of each run period. A drawing of the 1-A reactor is shown in Figure 1, and a photograph in Figure 2.

DRAWING OF I-A BATCH REACTOR

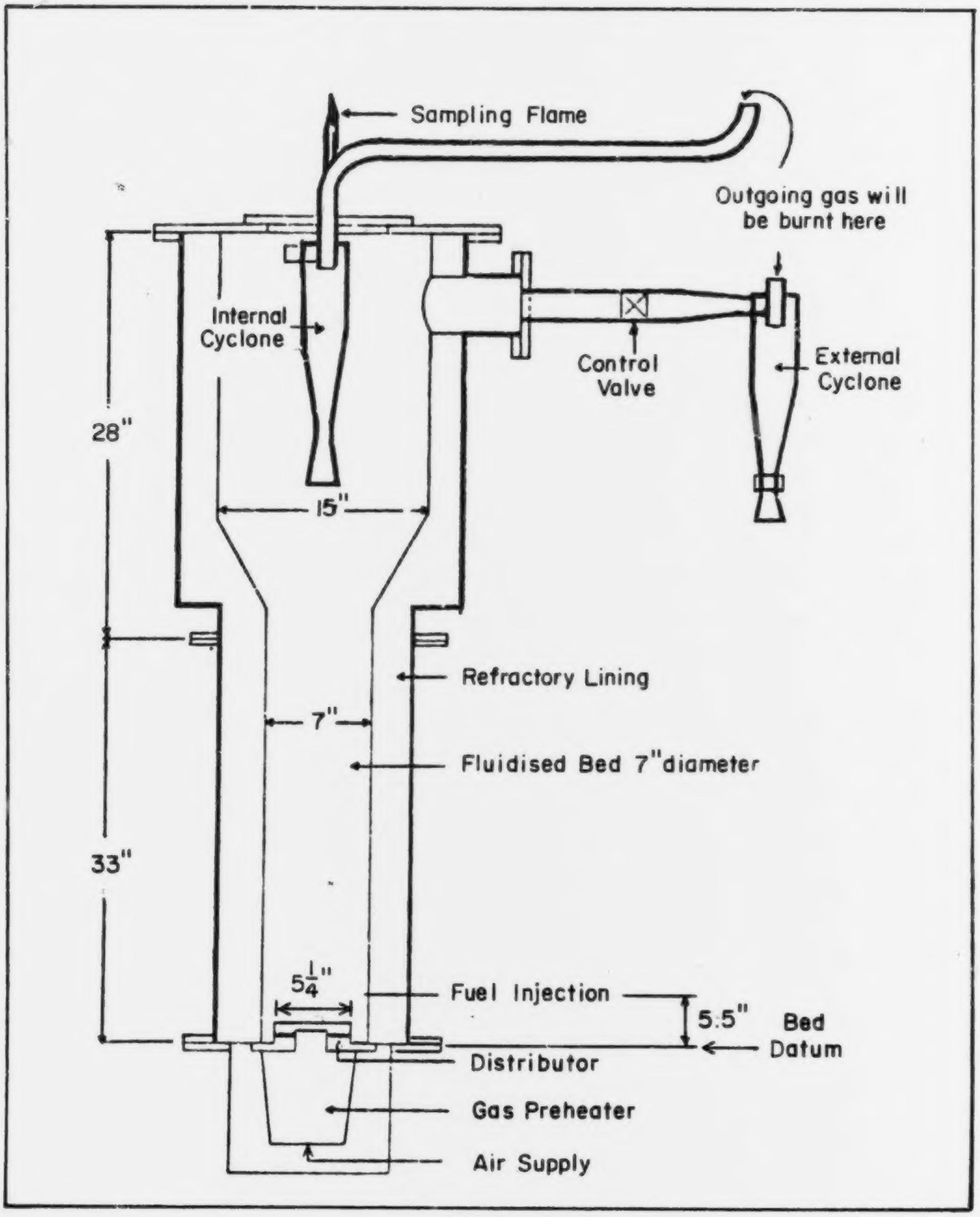

Sampling Flame
Outgoing gas will be burnt here
Internal Cyclone
Control Valve
External Cyclone
28"
15"
Refractory Lining
7"
Fluidised Bed 7" diameter
33"
$5\frac{1}{4}$"
Fuel Injection
5.5"
Bed Datum
Distributor
Gas Preheater
Air Supply

FIG.1.

PHOTOGRAPH OF I-A BATCH REACTOR

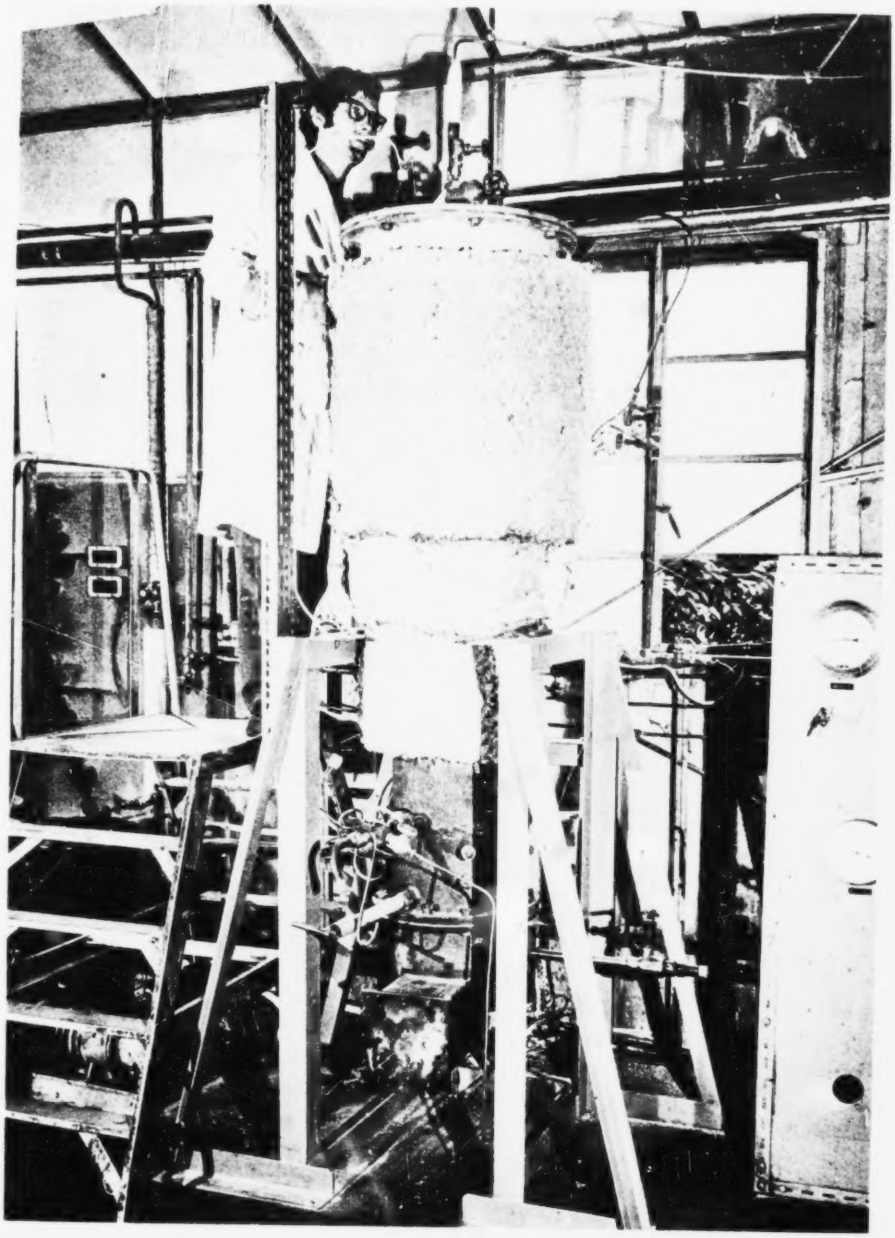


FIG. 2.

The reactor is made of a carbon steel shell lined with a high temperature castable refractory. The lower section, which contains the fluid bed, is 7 inches internal diameter, by 33 inches high. The upper section expands to 15 inches internal diameter and serves as a disengaging space for all but the finest particle size range of solids.

The distributor is a "top hat" design with a raised section 5 inches in diameter. It contains sixteen horizontal holes distributed around its circumference. Figure 3 is a sketch of the distributor.

BATCH UNIT DISTRIBUTOR

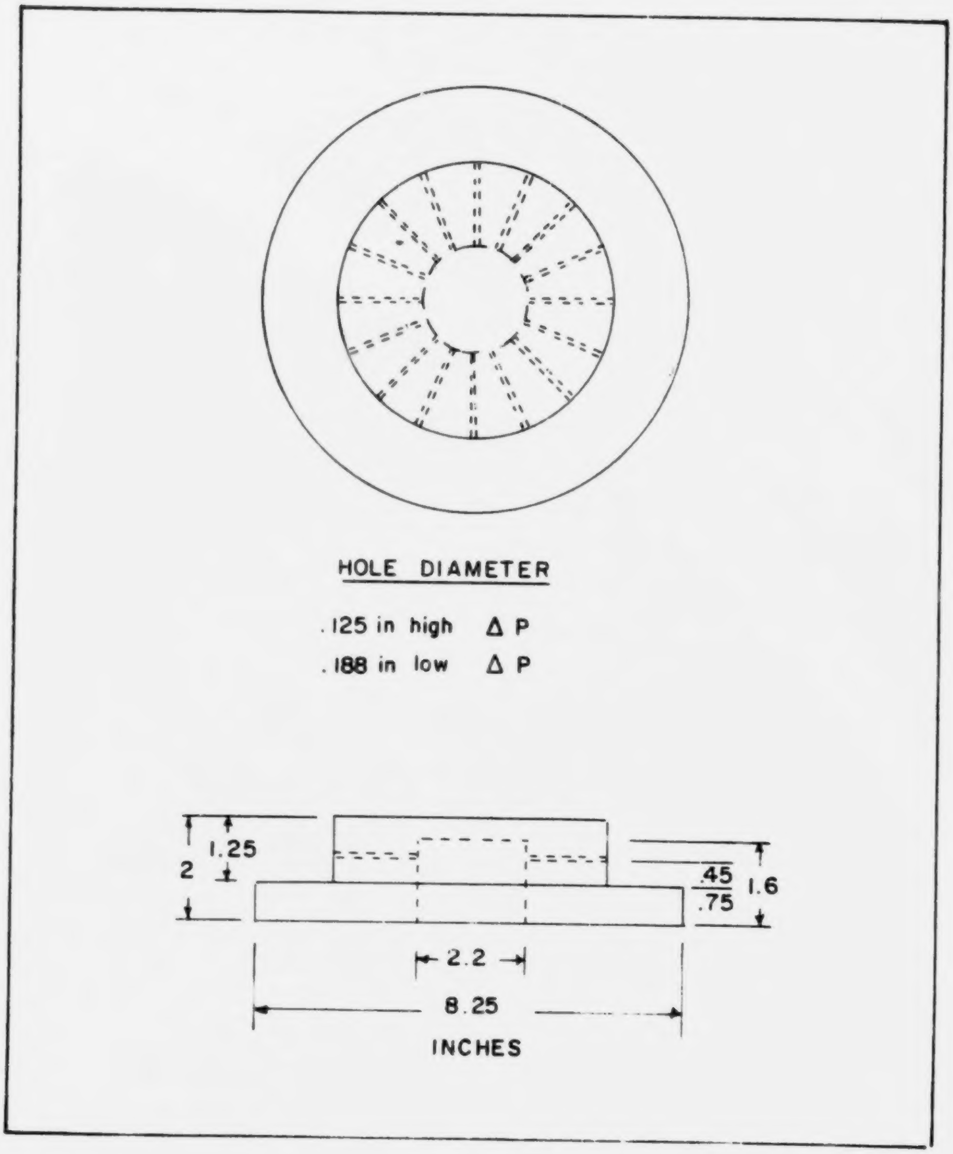

HOLE DIAMETER
.125 in high ΔP
.188 in low ΔP
1.25
2
.45
.75
1.6
2.2
8.25
INCHES

FIG. 3.

The 1-A unit has no provision for temperature control other than by adjusting the quantity of insulation around the reactor. This insulation was arranged to permit the unit to operate in the range of 800 to 900°C with air velocity of 4 ft/sec and air/fuel ratios in the range of 20 to 30% of stoichiometric. With constant air supply, the unit tends towards 900°C at the lower fuel rate and 800°C at the higher rate.

b. 4-A and 4-B Units

Two new batch reactors (designated 4-A and 4-B) were constructed specifically for this programme so that the variable study could proceed more rapidly. Their design was based to a large extent on that of the 1-A unit. Figure 4 is a drawing of the new reactor vessel with its various inlets and connections. The reactor is made of a carbon steel shell lined with a high temperature castable refactory.

CAFB BATCH UNIT NO. 4. REACTOR

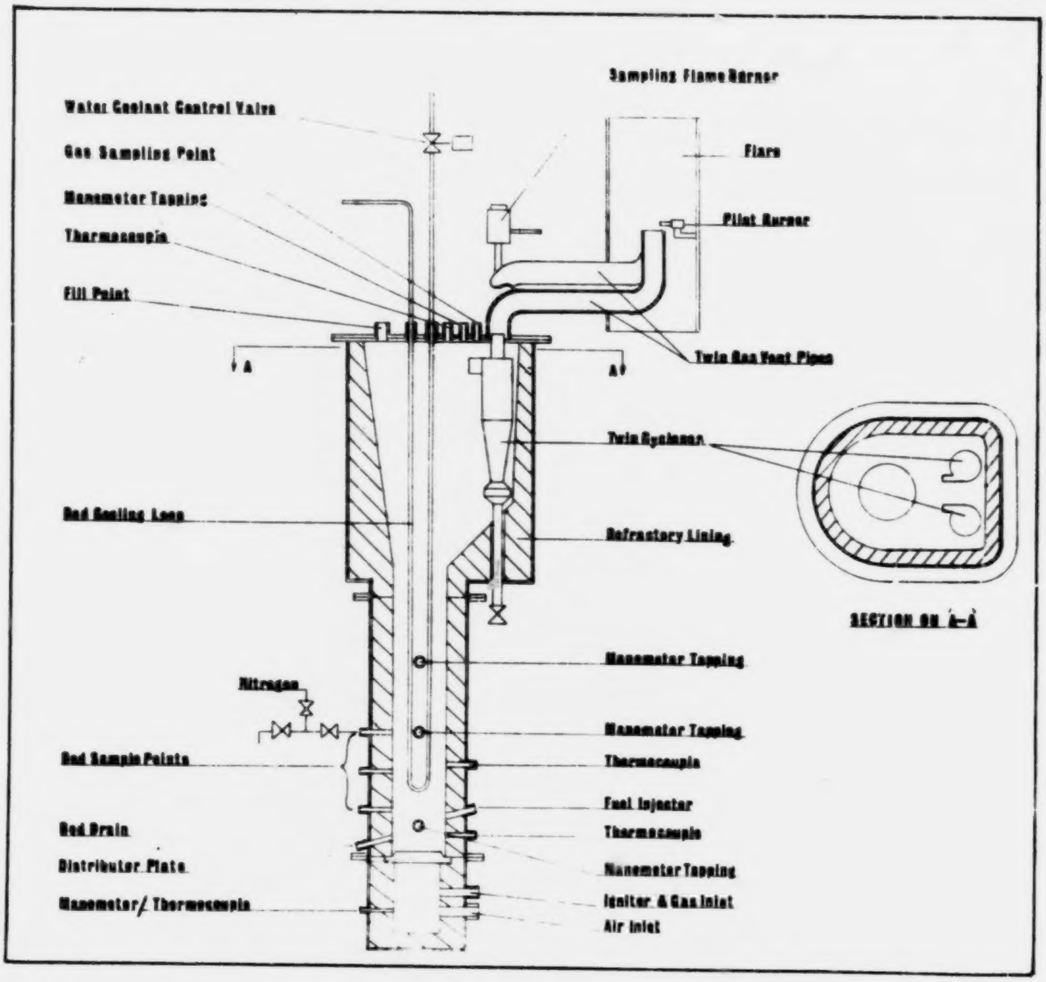

Sampling Flame Burner
Water Coolant Control Valve
Gas Sampling Point
Flare
Manometer Tapping
Pilot Burner
Thermocouple
Fill Point
Twin Gas Vent Pipes
A
A
Twin Cyclones
Bed Cooling Loop
Refractory Lining
SECTION ON A-A
Manometer Tapping
Nitrogen
Manometer Tapping
Bed Sample Points
Thermocouple
Fuel Injector
Bed Drain
Thermocouple
Distributor Plate
Manometer Tapping
Manometer/ Thermocouple
Igniter & Gas Inlet
Air Inlet

FIG. 4.

The lower section, which contains the fluid bed, is identical to that of the 1-A unit, being 7 inches in internal diameter by 33 inches high. Fuel oil enters the reactor through a single $\frac{1}{4}$ inch nozzle which protrudes one inch in from the reactor wall at a point $5\frac{1}{2}$ inches above the bottom.

The upper section was redesigned to permit installation of two internal cyclones and to allow drainage of the cyclones to external sample collectors during operation.

The distributor and plenum are the same as in the 1-A unit. The distributor is a top hat design with a raised section, 5 inches in diameter. It contains 16 horizontal holes distributed around its circumference.

A low pressure drop distributor with 3/16 inch diameter holes has been used for most of the work, but a higher pressure drop distributor with 1/8 inch holes has also been tested. Pressure drop characteristics of the two distributors are shown in Figure 5.

BATCH REACTOR

DISTRIBUTOR CHARACTERISTICS

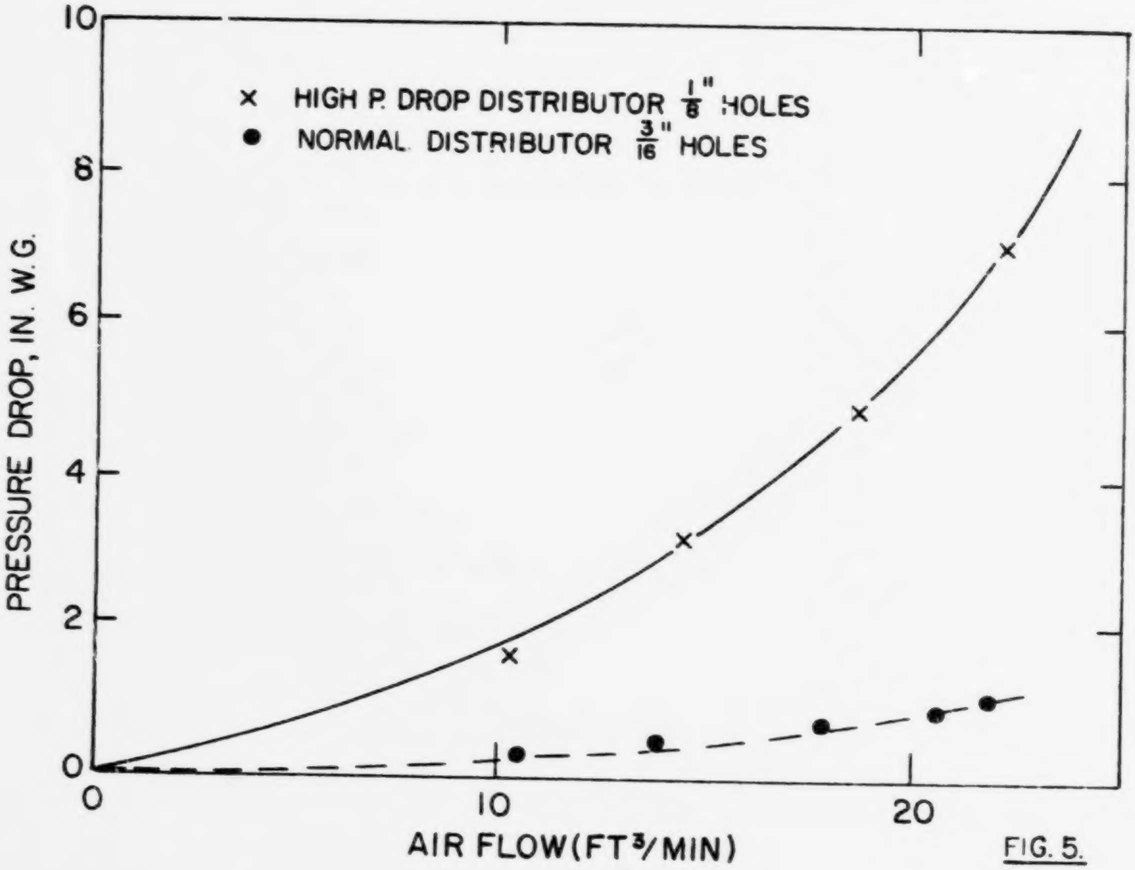


FIG. 5.

The 4-A and 4-B reactors contain internal cooling coils which allow reactor temperature to be controlled independently of fuel and air rates. This was not possible in the 1-A unit.

Figure 6 is a photograph of the new units with their control panel. These units have been designated No. 4-A (left hand unit facing panel) and No. 4-B.

PHOTOGRAPH OF NEW BATCH UNITS
4-A AND 4-B

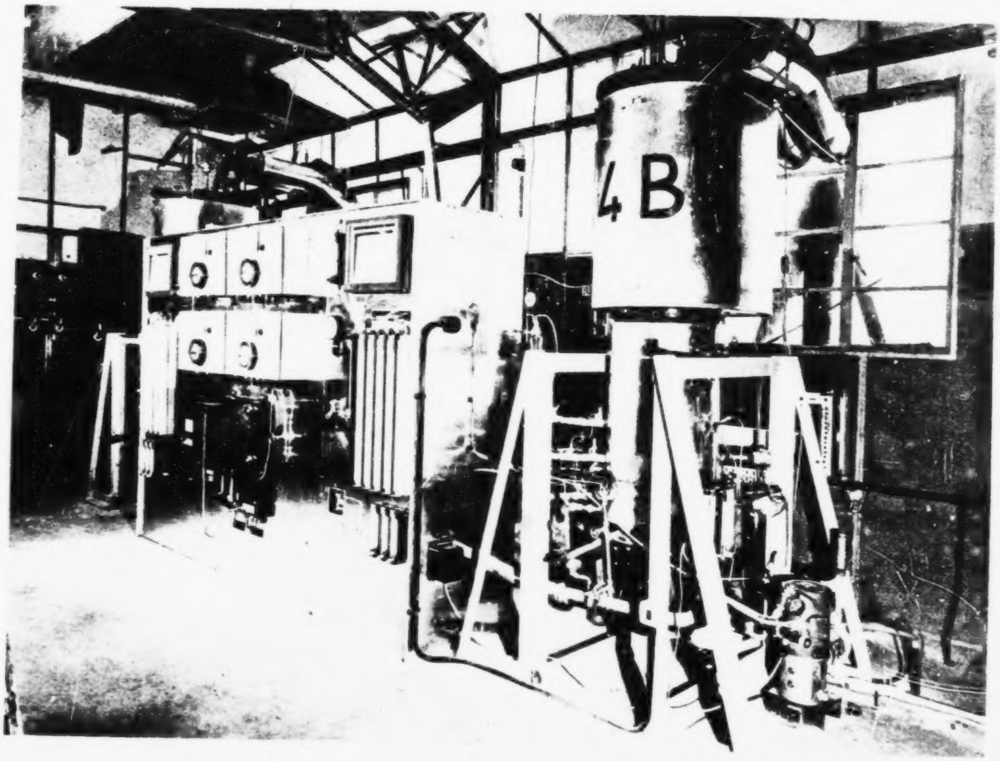


FIG 6

c. Tall Batch Unit

Several tests were performed in a third version of the batch units, designated as No-3 unit. The reactor section of this unit is the same diameter as that of the others, but the height is much greater; 17-ft compared with 3-ft in the straight sections of units 1-A, 4-A and 4-B. The No.3 unit also has an expanded section at its top, but it does contain an internal cyclone which returns fines to the bed. The No.3 unit had been constructed by Esso prior to beginning work on this contract.

d. Flow Plan

A flow plan of the batch units is given in Figure 7. This simple flow scheme applies both to the 1-A and to the 4-A and 4-B units.

FLOW PLAN OF CAFB BATCH UNIT

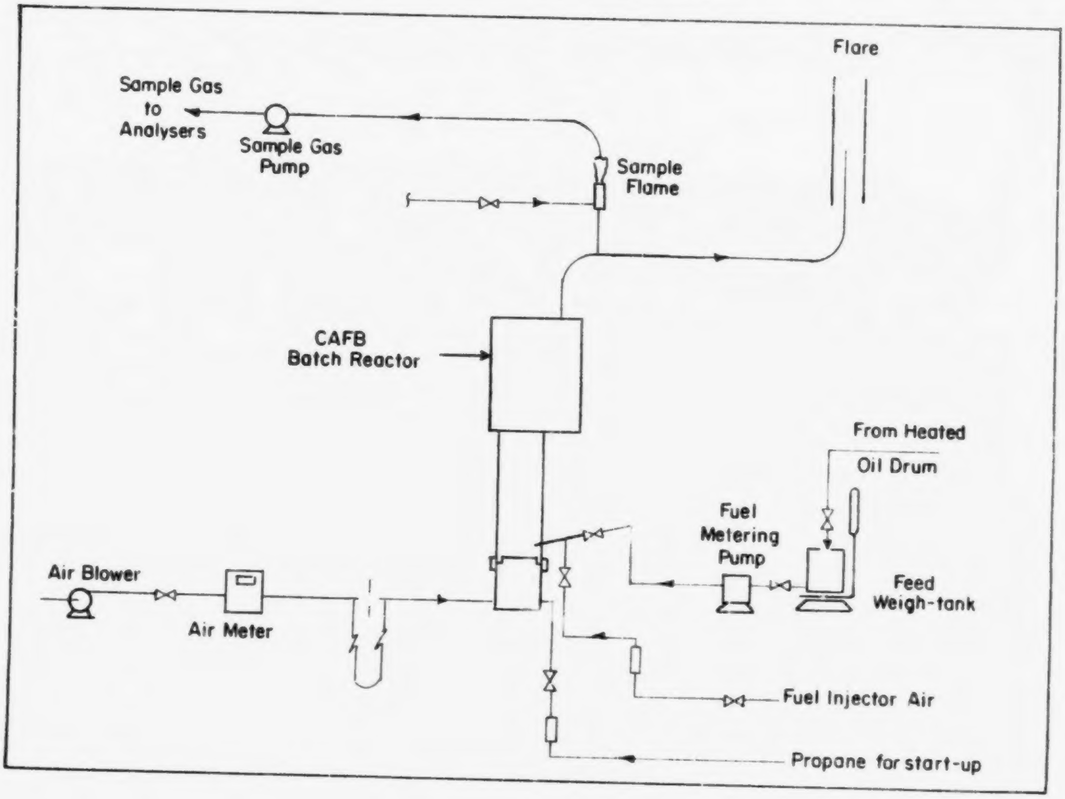

Flare
Sample Gas to Analysers
Sample Gas Pump
Sample Flame
CAFB Batch Reactor
From Heated Oil Drum
Fuel Metering Pump
Feed Weigh-tank
Air Blower
Air Meter
Fuel Injector Air
Propane for start-up

FIG. 7

e. Gas Analysis

Product gas leaves the unit through the cyclone outlet and passes to a flare outside the laboratory where it is burned. A portion of gas is burned in a sample flame located just above the reactor, and the combustion products are analysed for SO_2, O_2, CO and CO_2.

Most of the information obtained from the batch tests has been based on gas analysis. During gasification, the desulphurising efficiency of the bed has been calculated from carbon dioxide, oxygen and sulphur dioxide contents of the fully combusted product gas. Because of the wide range of sulphur compounds present in the gasifier product gas itself, this is the only practical method for measuring desulphurising efficiency. During regeneration, the exit gas was monitored for carbon dioxide, carbon monoxide, sulphur dioxide and oxygen.

The gas analysing equipment which was used in the tests is listed in Table I.

TABLE I

GAS ANALYSING EQUIPMENT

Analyser	Type	Manufacturer	Model	Response	Range
SO_2	Infra-red	Maihak	Unor 6	Continuous	0-1,000 ppm
SO_2	Infra-red	Maihak	Unor 6	"	0-20 % by vol.
SO_2	Conductimetric	Wosthoff	-	"	0-1000 ppm
CO_2	Infra-red	Maihak	Unor 6	"	0-20 % by vol.
CO	Infra-red	Maihak	Unor 6	"	0-20 % by vol.
O_2	Paramagnetic	Servomex	OA 137	"	0-25 % by vol.

2. Batch Unit Operating Procedures

Two types of study are carried out in the batch units; fresh bed tests, and cyclic tests. Differing procedures are used in the two studies.

a. Fresh Bed Tests

A new batch of calcined lime is used in each of the fresh bed gasification tests. These tests were used to screen rapidly the effects of a number of gasification variables such as bed depth, gas velocity, particle size, and air fuel ratio.

(1) Start-up

The bed was heated to operating temperature by combustion of propane (butane in the 1-A reactor) below the distributor and by direct injection of kerosine through the fuel nozzle. First of all, the bed space temperature was raised to 950°C by gas combustion below the distributor. Then, with gas and air to the unit shut off, approximately 9 lb of bed material was added. This was heated to 750°C using gas before switching to direct kerosine injection for the final 200°C rise. These latter steps were repeated until all the bed had been added and was at operating temperature. Although gas combustion could be used on its own to raise the bed to operating temperature, this route was avoided since it produced extremely high gas velocities through the distributor which caused appreciable attrition of the bed material.

(2) Calcination

The start-up procedure was also followed for calcination. Due to the strongly endothermic nature of the reaction, 950°C could only be attained when all the carbon dioxide was driven off. CO_2 emissions were also checked by gas analysis.

(3) Gasification

When the start-up procedure had been completed, the fluidising air was set to the desired level, the kerosine supply was terminated, and fuel oil was introduced at a predetermined feed rate. The sampling flame burner and external flare were lit and the gas analysers connected. Relevant data was recorded and bed samples taken from points 6 inches, 11 inches and 16 inches above the distributor at prescribed intervals. Bed sampling was carried out under nitrogen to prevent any oxidation of the sulphided lime. Samples of the gasifier product gas were also taken for separate analysis by gas chromatography.

(4) Regeneration

For regeneration, the fuel supply was switched off and air flow alone continued through the bed. The gas analysers were switched to the sampling system for regeneration and the relevant data were collected. A bed sample was collected after each regeneration.

(5) Shut-down

The bed temperature was allowed to drop to 700°C before the bed was removed through the drain point just above the distributor. Draining the unit is much easier when the bed is hot since the solids flow better under these conditions. When all the bed had been removed, the cyclone was emptied and all ancillary equipment shut off.

b. Batch Unit Cycle Tests

The cyclic tests are the nearest simulation to continuous gasifier operation that can be obtained in batch units. The same charge of lime is subjected to repeated cycles of sulphur absorption and regeneration. After each regeneration a portion of lime is removed and replaced by an equivalent amount of fresh limestone. The limestone calcines to lime during the early part of the next gasification cycle. Without replacement, the activity of the lime bed gradually declines. With replacement, the activity falls initially, but in a few cycles lines out at an equilibrium level which is influenced by the rate of replacement.

Enough cycles were performed at each set of operating conditions to establish the lined-out sulphur removal efficiency for those conditions.

Cyclic tests used the same calcination and start-up procedures as the fresh bed tests. However, after the initial start, a series of gasification and regeneration cycles followed each other. Sampling and gas analysis procedures also were the same as used in fresh bed tests.

When a regeneration cycle is complete, the fluidisation cools the bed very rapidly. Since solid sample withdrawal is very difficult in the absence of fluidisation, it was not possible to draw a regenerated bed lime sample before temperature dropped below the level needed for the next gasification cycle. Therefore the following procedure was adopted.

(1) When regeneration was complete as indicated by end of SO_2 emission and fall of bed temperature, fluidising air was stopped.

(2) Replacement limestone was added.

(3) Fluidising air was resumed and when temperature reached the desired point, oil feed was resumed.

(4) A solids sample was withdrawn to provide a measure of vanadium concentration between cycles and to return the bed to its proper level.

3. The Continuous CAFB Gasification Pilot Plant

A continuous CAFB gasifier pilot plant has been constructed at the Esso Research Centre to provide a demonstration of the CAFB process under continuous operating conditions and to provide a means for studying those operational variables which cannot be measured in batch reactors. Features of the pilot plant are summarised here. The design basis and construction details are discussed in greater detail in Appendix G.

a. Flow Plan and Layout

Figure 8 is a process flow plan of the continuous pilot plant. The heart of the system is the gasifier-regenerator unit cast of refractory concrete contained in an internally insulated steel shell. The product gas of the gasifier fires a 10 million Btu/hr pressurised water boiler. The hot water is heat exchanged with a secondary water circuit which loses its heat through a forced convection cooling tower. The rest of the system consists of the necessary blowers, pumps and instruments to operate the gasifier, regenerator, burner and solids circulating system.

CAFB PILOT PLANT FLOW PLAN

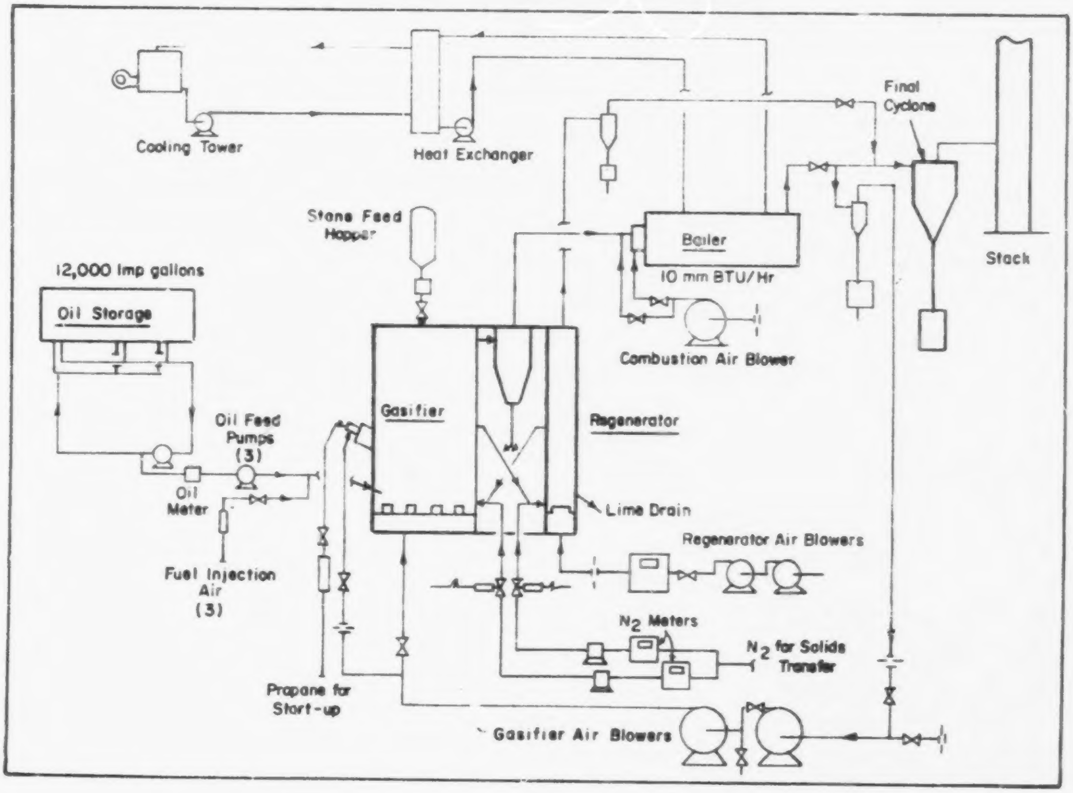

Cooling Tower
Heat Exchanger
Final Cyclone
Stack
Stone Feed Hopper
12,000 Imp gallons
Oil Storage
Boiler
10 mm BTU/Hr
Combustion Air Blower
Gasifier
Regenerator
Oil Feed Pumps (3)
Oil Meter
Lime Drain
Regenerator Air Blowers
Fuel Injection Air (3)
N_2 Meters
N_2 for Solids Transfer
Propane for Start-up
Gasifier Air Blowers

FIG.8.

Figure 9 shows the layout of the pilot plant equipment within its building. The gasifier itself sits within a pit to permit alignment of the gasifier outlet duct with the burner inlet. Fuel pumps, flow meters, and start up burner controls are mounted on a mechanical equipment console at the pit side. Electrical instrumentation and manometers are mounted in a separate control cabinet. Gasifier blowers are located in a separate blower house outside the main building, and the cooling tower is mounted on the roof.

GENERAL PLANT LAYOUT

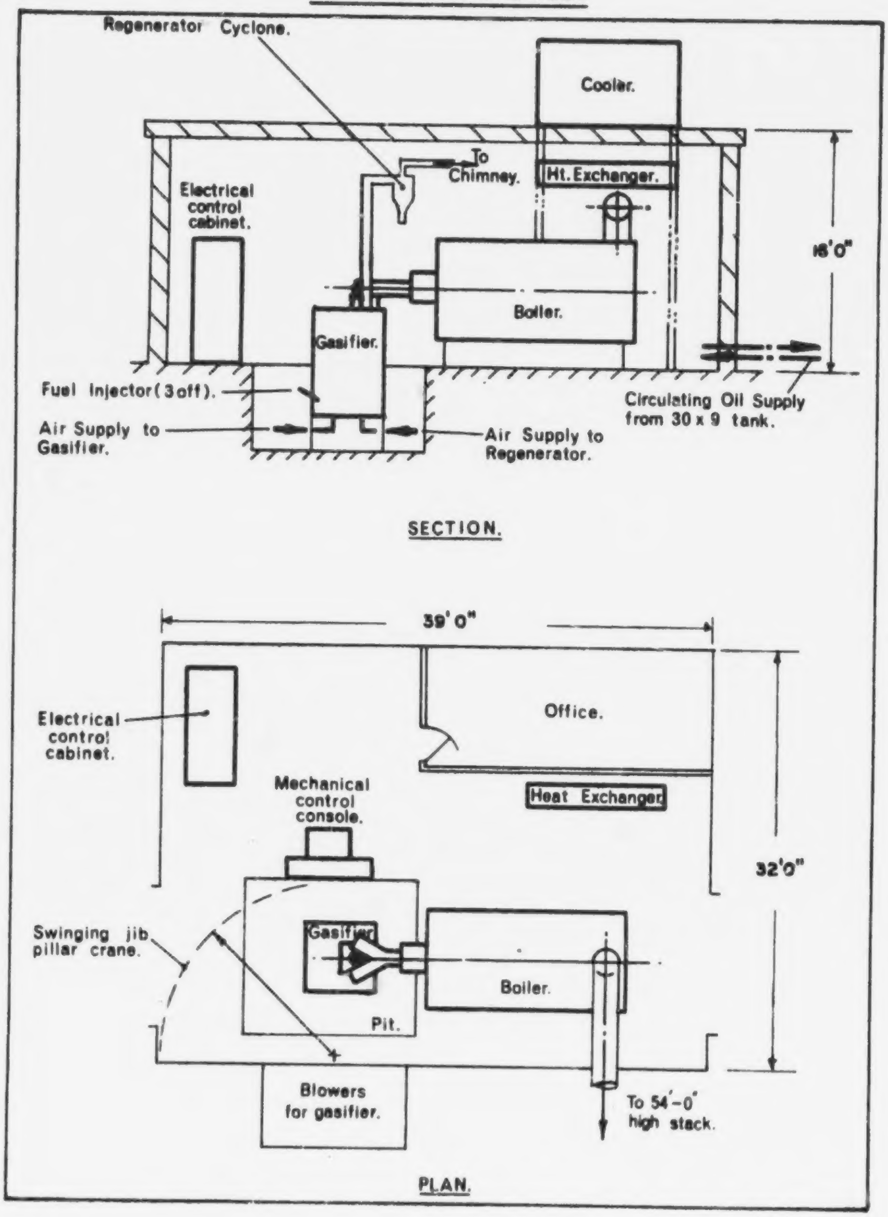

Regenerator Cyclone.
Cooler.
To Chimney.
Ht. Exchanger.
Electrical control cabinet.
18'0"
Boiler.
Gasifier.
Fuel Injector (3 off).
Air Supply to Gasifier.
Air Supply to Regenerator.
Circulating Oil Supply from 30 x 9 tank.
SECTION.
39'0"
Electrical control cabinet.
Office.
Heat Exchanger
Mechanical control console.
32'0"
Swinging jib pillar crane.
Gasifier
Boiler.
Pit.
Blowers for gasifier.
To 54'-0" high stack.
PLAN.

FIG. 9.

Figure 10 is an interior view of the completed pilot plant showing gasifier unit, boiler and mechanical equipment console.

CAFB PILOT PLANT INTERIOR

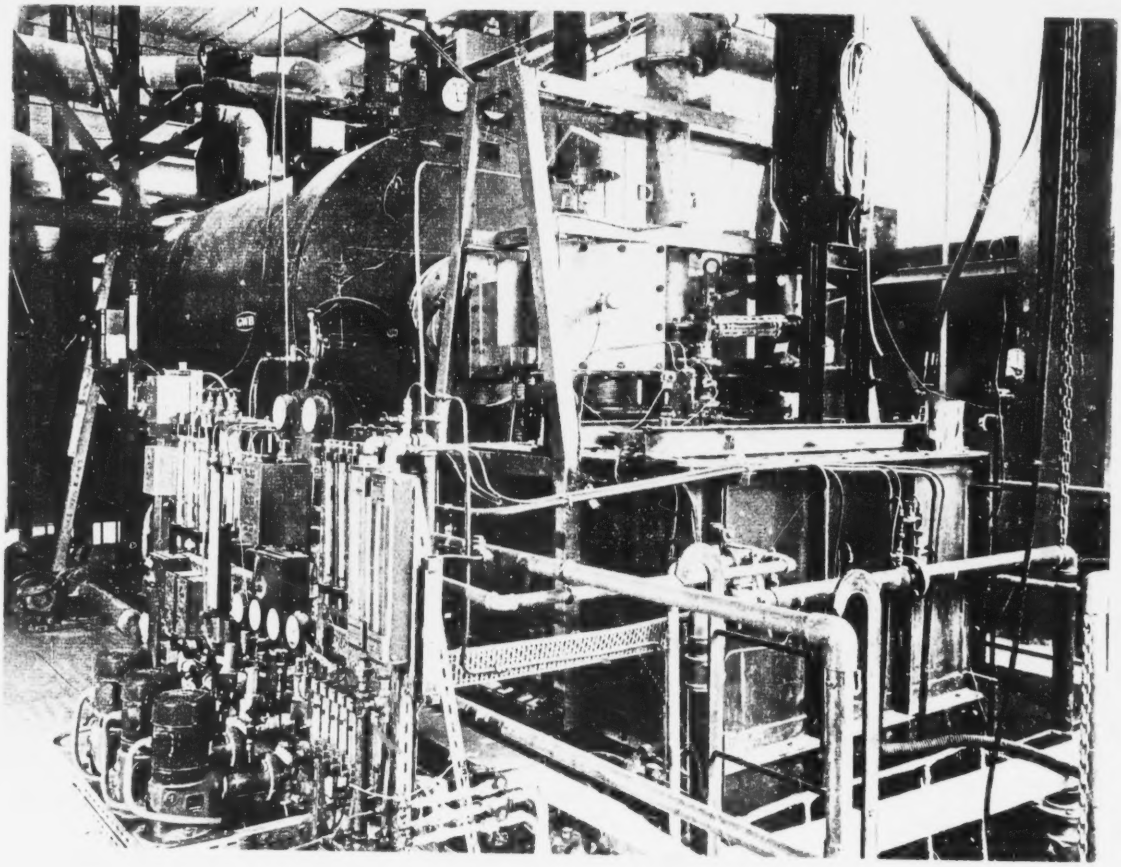


FIG. 10.

Figure 11 shows the pilot plant exterior with oil feed storage tank, rear end of boiler, cooling tower, and flue stack.

CAFB PILOT PLANT EXTERIOR

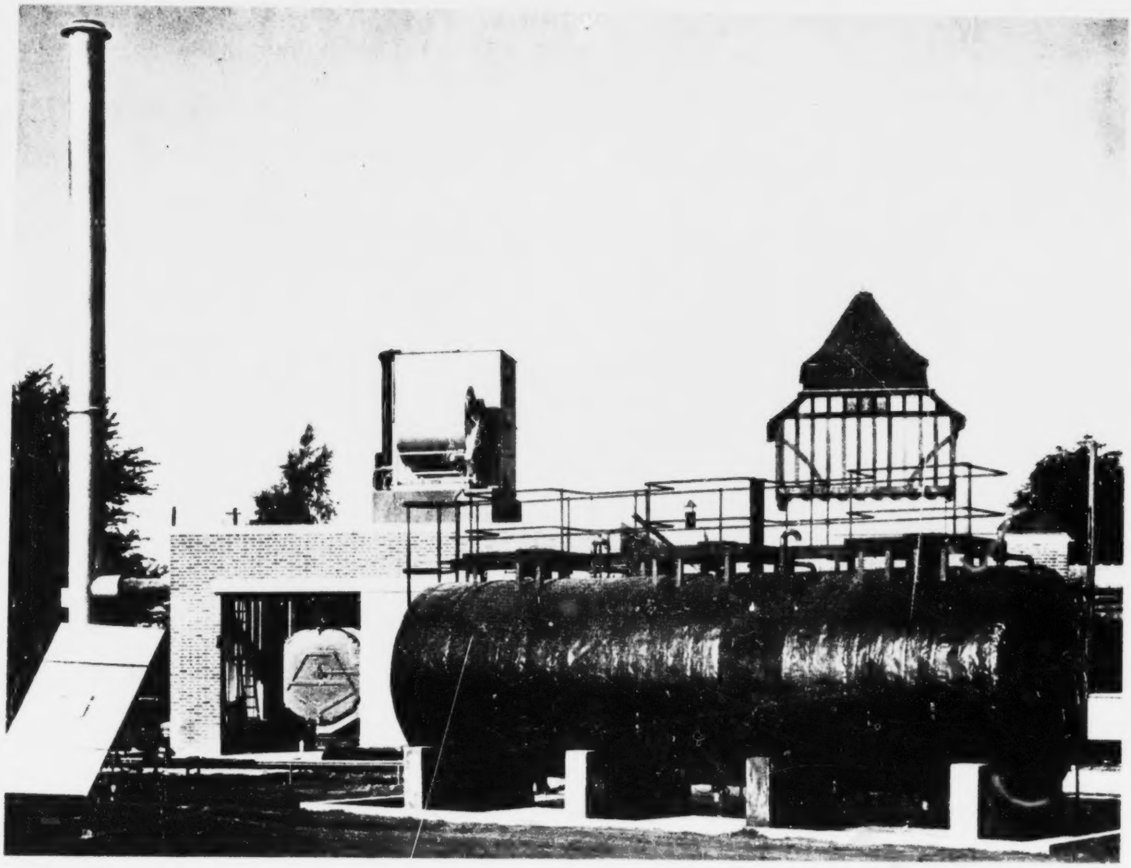
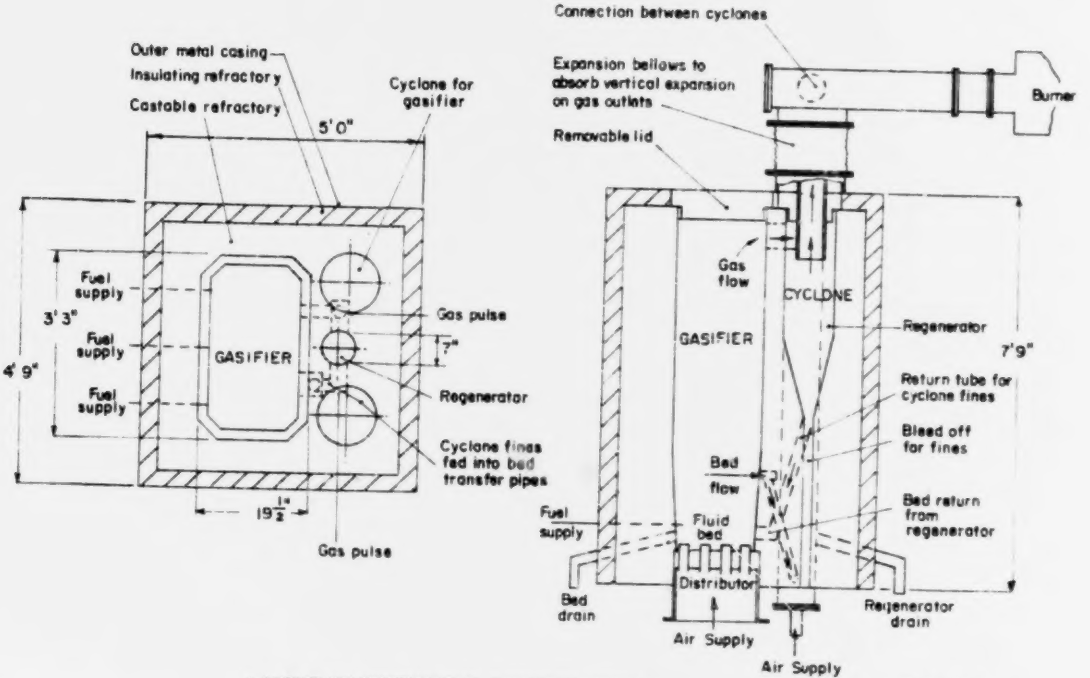


FIG. 11

b. Gasifier-Regenerator Unit

The gasifier and regenerator reactors are cavities in a single refractory concrete block. The block contains other cavities which make up the gasifier outlet cyclones, the gas transfer ducts, and the transfer lines through which solids circulate between gasifier and regenerator. Figure 12 is a drawing of this gasifier-regenerator assembly. The gasifier cavity is rectangular in cross section, tapering from 17.5 x 37 inches at the distributor level to 19.5 x 39 inches at the 21 inch level. The upper portion has parallel sides. The regenerator is 7 inches in diameter throughout.


Outer metal casing
Insulating refractory
Castable refractory
5'0"
Cyclone for gasifier
Fuel supply
Fuel supply
Fuel supply
3'3"
4'9"
GASIFIER
Gas pulse
7"
Regenerator
Cyclone fines fed into bed transfer pipes
19 1/2"
Gas pulse
Connection between cyclones
Expansion bellows to absorb vertical expansion on gas outlets
Burner
Removable lid
Gas flow
CYCLONE
GASIFIER
Regenerator
7'9"
Return tube for cyclone fines
Bleed off for fines
Bed flow
Fuel supply
Fluid bed
Bed return from regenerator
Distributor
Bed drain
Air Supply
Air Supply
Regenerator drain

LAYOUT OF CONTINUOUS GASIFIER UNIT

FIG. 12.

The gasifier air distributor consists of 32 nozzles each of which contains six horizontal holes of 0.177 inch diameter. The nozzles are stainless steel mounted on a refractory covered carbon steel plenum.

The regenerator distributor is of the same design as that used in the batch reactors, a refractory top-hat shape with sixteen horizontal holes. Hole diameter in the continuous unit distributor, however, is 0.25 inch.

Assembly of the gasifier unit steel framework and casting of the high density refractory concrete was performed with the gasifier in place at the Esso Research Centre. After pouring each lift of concrete, it was vibrated to provide maximum density.

Figure 13 is a photograph showing the gasifier at an early stage of construction. The plastic patterns which form the regenerator and solids transfer ducts are in place together with the wooden pattern of the lower gasifier bed. Three of the pre-insulated mild steel encasing walls also are erected.

GASIFIER DURING CONSTRUCTION

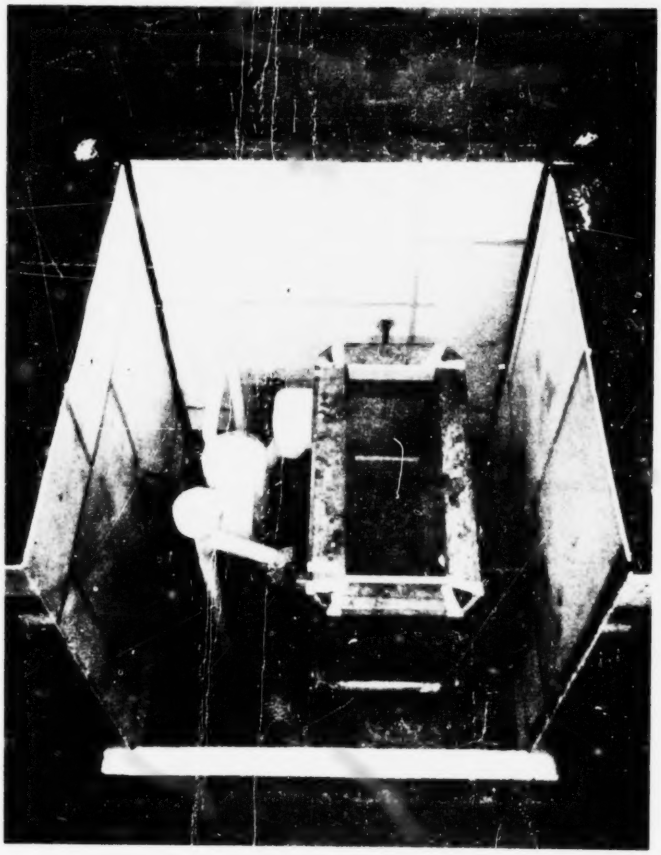


FIG. 13.

c. Gasified Fuel Burner

The standard oil burner of the 10 million Btu/hr boiler was replaced with an experimental burner designed to handle the hot gasifier product.

A general arrangement of the gasified fuel burner is shown in Figure 14. It will be seen that the air required for complete combustion is introduced in two stages. Roughly about 10% of the air enters a crude injector and is pre-mixed with the gas. The remainder of the air is fed tangentially into a swirl chamber and emerges from an annular nozzle which is concentric with the gas nozzle. The design of this burner was purely empirical and was based on recommended pressure drops at the nozzles. The main gas duct is sized to give a flow velocity of about 60 ft/second, the gas nozzle gives a pressure drop in the region of 3" w.g., and the air nozzle gives a pressure drop of about 4-5" w.g., all under full load conditions described in Appendix G.

MAIN GAS BURNER

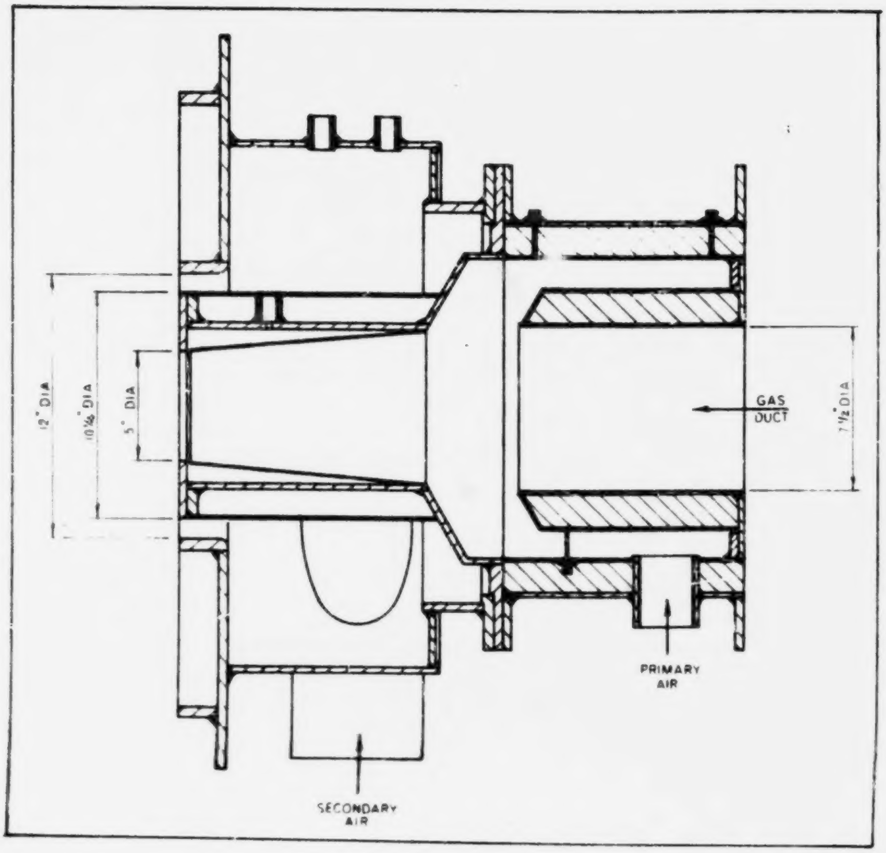


FIG. 14.

A similar burner of about 1/20th of the output was tested on a batch unit prior to the construction of the large burner and was found to give a satisfactory performance. The remarkably good behaviour of the large burner, which required no modifications despite the 20/1 scale up, suggests that the gas is of excellent combustion quality.

A pilot light fired by propane gas is used for lighting purposes. This pilot required a certain amount of development work since a spoiler had to be fitted to prevent its extinction by the secondary air blast. The whole burner system has been found to be reliable in operation.

d. The Fuel System

The burner which is used to heat the gasifier to its working temperature is fuelled by propane gas. It is fitted to an end wall of the gasifier and is angled downwards so that the hot gases impinge on the fluidised bed. A standard commercial burner of about 700,000 Btu/hr maximum firing rate is employed.

Three fuel-oil injectors are provided for use during gasifying operation and these penetrate the side wall of the gasifier opposite the cyclones, as shown in Figure 12. These injectors are simple $\frac{1}{2}$" bore tubes made of a heat resistant iron-chrome alloy. In the first continuous gasifier run the full diameter injector was used. In the second run a restriction orifice drilled with a 3/16 inch hole was placed in each injector tip to increase injection velocity. In the third run, these inserts were drilled to $\frac{1}{4}$ inch diameter. The injectors are positioned approximately $4\frac{1}{2}$" above the top surface of the distributor and project into the bed about 3". Some air is injected through these tubes with the fuel, to aid injection and prevent coking in the hot tube ends. Each injector has its own fuel oil metering pump drawing its oil supply from a heated circulating main. It is also possible to switch to an alternative distillate fuel supply which is used during the start-up period in order to reduce the propane consumption.

e. The Stone Handling System

Standard commercial equipment, including an industrial vacuum cleaner, is used for handling the stone outside the gasifier. The stone feed hopper is mounted upon the lid of the gasifier, and a vibrator is used to control the rate at which the stone enters the gas space. The hopper is pressurised by a nitrogen bleed during operation in order to prevent the escape of gas from the unit and the consequent deposition of tars in the feed line. This system has not proved to be entirely satisfactory. A more positive feed control would be preferred and modifications are planned.

f. Safety Provisions

The installation is provided with a comprehensive alarm system which detects and indicates faults occurring in the boiler and its cooling system, the gasifier and the experimental burner. In some instances, such as for example a flame-out in the boiler, or failure of the pilot light, shutdown is automatic. Details of the alarm system of the various alarm actions are given in Appendix H.

g. Process Controls

Figure 15 is a diagram of the pilot plant instrumentation system. Automatic control boxes are used to regulate regenerator temperature, regenerator bed level, and gasifier bed level. A packaged pressurisation system maintains constant boiler cooling water pressure and temperature. All other systems are manually controlled by the process operator. Dashed lines in Figure 15 show the indicators and control valves normally used by the operator. Manometers indicate pressures and pressure differences in most applications. In four instances pressure differences are also detected by pneumatic delta pressure cells and transmitted to recorders. Pressure switches also are employed in several locations to operate warning lights for abnormal conditions.

CAFB PILOT PLANT INSTRUMENT FLOW PLAN

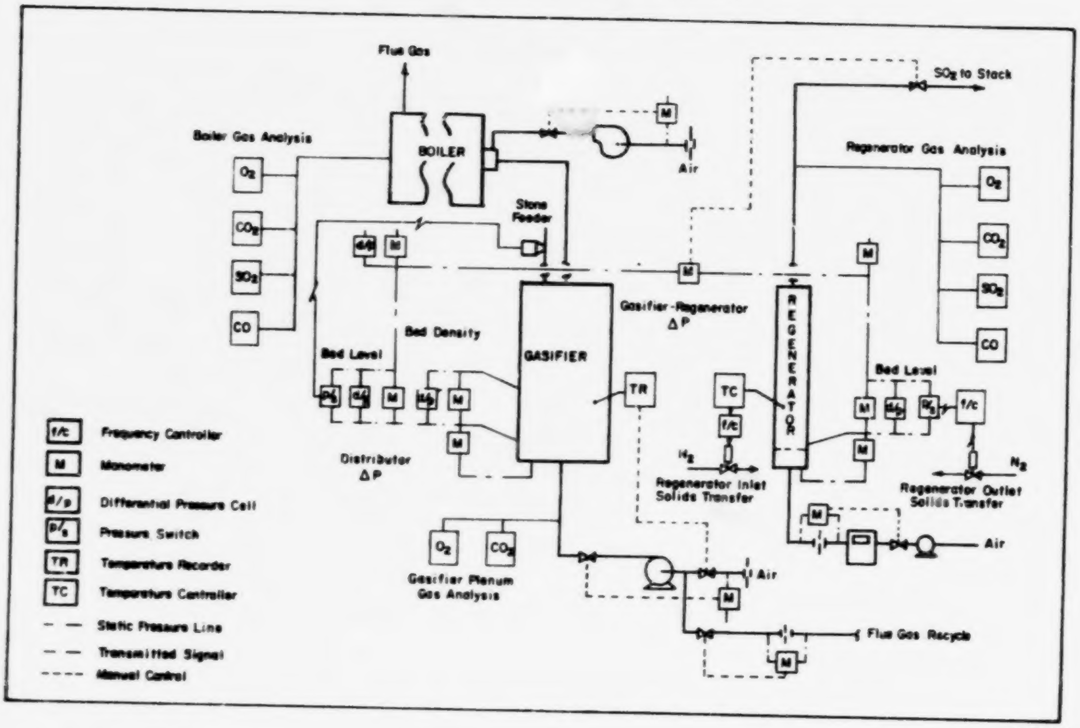

Flue Gas
Boiler Gas Analysis
O_2
CO_2
SO_2
CO
BOILER
M
Air
Stone Feeder
d/p
M
SO_2 to Stack
Regenerator Gas Analysis
O_2
CO_2
SO_2
CO
M
Gasifier-Regenerator ΔP
REGENERATOR
GASIFIER
Bed Density
Bed Level
P/s
d/p
M
d/p
M
M
Distributor ΔP
TR
TC
f/c
H_2
Regenerator Inlet Solids Transfer
Bed Level
M
d/p
P/s
f/c
M
N_2
Regenerator Outlet Solids Transfer
M
Air
O_2
CO_2
Gasifier Plenum Gas Analysis
Air
M
Flue Gas Recycle
M
f/c Frequency Controller
M Manometer
d/p Differential Pressure Cell
P/s Pressure Switch
TR Temperature Recorder
TC Temperature Controller
Static Pressure Line
Transmitted Signal
Manual Control

FIG. 15.

h. Analysers

On stream analysers monitor the compositions of the key gas streams. Table II lists these analysers and their applications.

Table II

CAFB Pilot Plant Gas Analysers

Gas Stream	Component	Analyser	Operating Principle	Range
Air-Flue Gas Mix to Gasifier Plenum	O_2	Servomex OP 250	Paramagnetic	0 - 25% by vol
	CO_2	Maihak Unor 6	Infra Red	0 - 10% by vol
Boiler Flue Gas Sampled at fire tube outlet	O_2	Servomex OA 137	Paramagnetic	0 - 5% by vol
	CO_2	Maihak Unor 6	Infra Red	0 - 20% by vol
	CO	Maihak Unor 6	Infra Red	0 - 20% by vol
	SO_2	Maihak Unor 6	Infra Red	0 - 1000 ppm
	SO_2	Wösthoff	Electrical conductivity of H_2O_2 - SO_2 reactor products in solution	0 - 1000 ppm
Regenerator	O_2	Servomex OA 137	Paramagnetic	0 - 2.5% by vol
	CO_2	Maihak Unor 6	Infra Red	0 - 10% by vol
	SO_2	Maihak Unor 6	Infra Red	0 - 20% by vol

A sampling pump and filter system supplies each gas stream to its appropriate group of analysers. A multipoint instrument records outputs of all instruments except the Wösthoff SO_2 analyser which has its own continuous recorder.

i. Operating Procedures

Details of continuous unit operating procedures for start-up, operation, decoking and shutdown are given in Appendix I.

4. Cold Model of Gasifier

Before design of the continuous gasifier-regenerator unit was finalised, a steel and plastic mock-up of approximately the same dimensions was constructed. This unit was used for testing the bed transfer system to determine its capacity and to determine the effectiveness of alternate arrangements of the transfer ducts. A photograph of the cold unit appears in Appendix Figure G-4.

5. Materials

Table III lists properties of the fuel oils and limestones used in these CAFB studies. Basically there have been three stones and two oils. However, different shipments from the same oil or stone source have had minor variations. These differences are also indicated in the table.

a. Limestone

The U.K. limestone used in early CAFB studies is a relatively high purity material quarried in Denbighshire in North Wales. When calcined it gives a lime containing 97.5% CaO. This stone was used as a basis for comparison of the performance of the U.S. limestones. It also was the material used in the continuous pilot plant operation trials.

Two U.S. limestones were supplied by NAPCA (now OAP) for the limestone comparison and selection study. These stones were shipped in steel drums and crushed and sieved at the Esso Research Centre. Both U.S. stones were of lower calcium oxide content than the U.K. stone and represent classes of material more widely available in the United States than the high purity stone. The stone designated 1690 in the Bituminous Coal Research Inc (BCR) listing comes from Ogdensburg, New York. It has the lowest calcium content of the three stones tested and is high in both magnesia and silica. The stone BCR 1691, from Seneca Falls, New York, also is high in silica but is low in magnesia. It is intermediate in calcium content and is similar in composition to Pennsylvania "cement rock".

After the preliminary comparison study, a larger supply of BCR 1691 was obtained for more detailed variable studies. This main supply had less silica and more CaO than the original sample, but in other respects was quite similar in composition.

b. Oil

The fuel oil used in early CAFB studies at the Esso Research Centre, Abingdon, was a blend of Middle Eastern and Venezuelan stocks. This had been chosen for the early tests because it was intermediate in both sulphur and vanadium concentrations and because it was available in large supply at the Research Centre. The initial sample supplied by NAPCA (now OAP) was a Venezuelan heavy fuel oil from the Creole Petroleum Co. Amuay refinery. When tests confirmed the suitability of this fuel, a larger supply was obtained in drums for the variable study. A quantity of oil from this same source in storage in the U.K. was used for runs in the continuous pilot plant. Table III lists inspection results of the oil samples and shows the minor variations between the different Venezuelan shipments. The most significant difference appears to be the lower vanadium content of the initial one drum sample. The initial U.K. test fuel reflects the high sulphur and lower vanadium content which is characteristic of its Middle Eastern component.

TABLE III

Composition of CAFB Test Limestones and Fuel Oils

		Limestone Composition			
Stone:		Denbighshire	BCR 1690	BCR 1691 (a)	BCR 1691 (b)
Component					
CaO	wt. %	55.2	26.7	45.0	49.1
MgO	"	0.27	17.5	2.02	2.08
SiO_2	"	0.32	13.5	14.2	8.44
Fe_2O_3	"	0.07	0.83	0.53	0.47
Al_2O_3	"	0.06	5.10	1.14	1.42
Na_2O	"	0.03	0.19	0.17	0.14
CO_2 released	"	43.4	35.5	35.4	37.0
S as SO_3	"	-	0.59	1.0	0.85
Vanadium ppm	"	-	-	< 10	< 10

	Fuel Oil Properties			
Oil	Venezuelan, Amuay (c)	Venezuelan, Amuay (d)	Venezuelan, Amuay (e)	U.K.
Property				
Specific gravity	.95	.957	.955	.95
Kinematic viscosity				
Cs at 140°F	-	201	221	98.2
180°F	54.7	(75) interpolated		-
210°F	-	41.4	39.3	26.2
Carbon % by wt.	85.9	85.9	85.6	84.8
Hydrogen % by wt.	11.5	11.3	11.4	11.6
Sulphur " "	2.22	2.35	2.48	3.03
Nitrogen " "	0.41	0.35	0.26	-
Conradson Carbon % by wt.	7.5	11.6	10.9	9.32
Asphaltenes % by wt.	6.3	7.1	4.8	3.72
Vanadium ppm	270	366	345	140
Nickel "	-	43	40	20
Sodium "	13	36	35	37
Iron "	4	3	4	1

(a) Initial sample of BCR 1691 used in screening tests.
(b) Larger supply of BCR 1691 used in variable studies.
(c) Initial sample of oil used in screening tests.
(d) Larger supply of oil used in variable studies.
(e) Large supply of oil used in pilot plant runs.

6. Experimental Programme

The experimental programme of Phase I of this CAFB project has included the following studies:

a Batch reactor oil and lime screening

b Batch reactor fresh bed variable studies

c Batch reactor cyclic test variable studies

d Continuous gasifier commissioning and operating trials.

Table IV summarises test conditions used in these separate studies.

Table IV

Summary of Phase I CAFB Experimental Conditions

Programme Item	Batch Unit Screening	Batch Unit Fresh Bed Variables	Batch Unit Cyclic Variables	Pilot Plant Operation
Number of Tests	121	33	10 tests, 246 cycles	3
Test Duration	60 to 300 min	80 to 200 min	35 to 68 min absorption cycles	91 to 202 hours
Limestone Used	U.K., Denbighshire; U.S., 1690; U.S., 1691	U.S., BCR 1691	U.S., BCR 1691; U.K., Denbighshire	U.K., Denbighshire
Oils Used	U.K., 3% S; Venezuelan, 2.2% S	Venezuelan 2.3% S.	Venezuelan 2.3% S	Venezuelan 2.5% S
Gasifier Temperatures	740°C to 900°C	845°C to 950°C	800°C to 880°C	820°C to 920°C
Regenerator Temperatures (peak)	1000°C to 1090°C	1000°C to 1140°C	955°C to 1080°C	1050°C to 1100°C
Gasifier Bed Depth (static)	15.2 in to 18.1 in	10 in to 20 in.	12.7 in to 26.7 in	13 in to 23 in.
Gasifier Superficial Velocity	3.5 to 4.3 ft/sec	4.3 to 7.9 ft/sec	3.7 to 6.9 ft/sec.	2.8 to 4.3 ft/sec.
Solids Particle Size Range	800/1200 microns; 600/1400 microns	600/1400; 1200/3175; 300/3175 microns.	600/1400; 300/3175 microns	300/3200- 800/3200 microns
Lime Replacement Rate	0.43 to 0.87 mole CaO/mole S	-	0.76 to 2.9 Mole CaO/mole S	0.54 to 4.8 mole CaO/mole S
Air/Fuel Ratio	17.7 to 27% of stoichiometric	14.8 to 38.5% of stoichiometric	21.2 to 38.8% of stoichiometric	15 to 31% of stoichiometric
Oil Feed Rate	74 to 121 lb/hr-ft^2	89.5 to 178 lb/hr-ft^2	51 to 138 lb/hr-ft^2	61 to 82 lb/hr-ft^2

C Results and Discussion

This Phase I CAFB study has consisted of four activities - each with its own programme of experiments and apparatus.

These activities are treated separately in the Appendix to this report. However, in this section, the discussion is organised around the CAFB variables and their effects. An attempt is made to summarise and interpret the findings in light of the total study rather than in sequence of the particular experiments which produced the information.

1. Oil and Limestone Comparisons

Tests to define the suitability for CAFB of oils and limestones available in the United States were performed early in the programme, so that larger supplies of the appropriate materials could be obtained for detailed variable studies.

a. Oils

One oil sample had been supplied as being typical of a heavy fuel oil available on the East coast of the United States. This Venezuelan oil contained a much higher vanadium to sulphur ratio (270 ppm/2.22%S = 121) than the blend used in prior CAFB experiments (140 ppm /3.03%S = 46). Thus for an equal sulphur pickup, lime would be exposed to a much higher vanadium level with the Venezuelan fuel. The test results in Figure 16 show nearly equal sulphur removal performances with the two oils in fresh bed batch tests using UK lime in the 1-A test unit. Both gave better than 90% sulphur removal at calcium utilisations approaching 10%. On the basis of this comparison, the Venezuelan oil was judged to be satisfactory for further study and a larger supply was obtained. All subsequent work, including runs in the continuous gasifier pilot plant, has confirmed that high levels of sulphur removal can be obtained from this oil in spite of its high vanadium content.

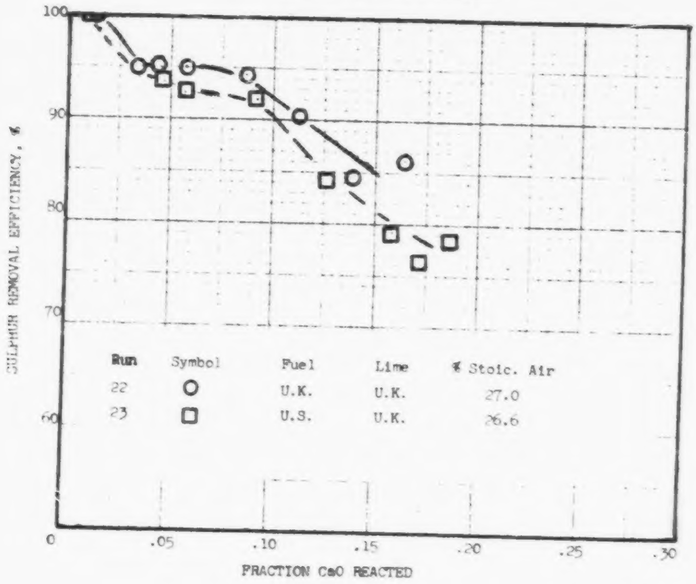

SULPHUR REMOVAL COMPARISON FOR U.S. AND U.K. FUELS

FIG.16.

b. Limestone Comparison Tests

Samples of two limestones available in the North-eastern United States were supplied. These stones were compared with each other and with the UK stone in a series of cyclic tests in the 1-A batch unit. The UK stone was subjected to 21 cycles of gasification-regeneration at conditions listed below.

TABLE V

Conditions for Limestone Comparison

Limestone Particle Size	850 - 1200 microns
Gasification Bed Temperature	800 - 900 °C
% Stoichiometric Air for Gasification	25
Superficial Air Velocity	4 ft/sec
Bed Depth (static)	15.5 in
Gasification Cycle Length	60 min
Maximum Regeneration Temp	1100 °C
Lime Replacement Rate (as limestone)	0.75 wt CaO/wt S.

Figure 17 shows the results - Sulphur removal efficiency declined for nearly ten cycles and then lined out at approximately 72%. This result agrees well with the curve of lime replacement versus sulphur removal efficiency established earlier in other apparatus (Appendix B, Figure B - 7). The data on which Figure 17 is based are listed in Tables C - 1 and C - 3, runs 33 - 53.

BED LIFE STUDY

(U.K. FUEL AND STONE)

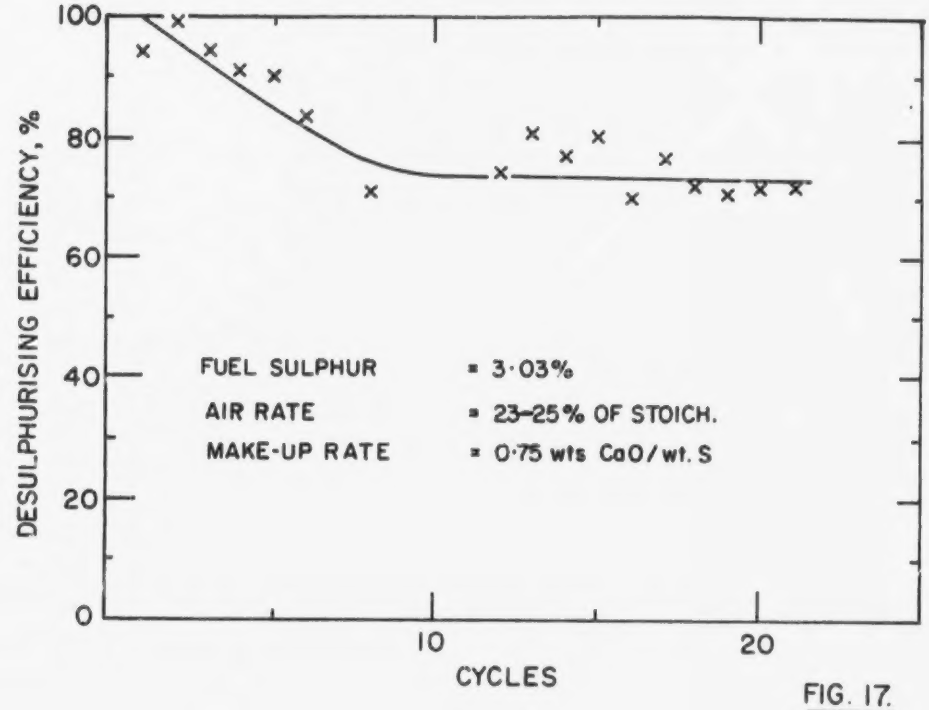


FIG. 17.

Stone BCR 1691 was tested under nearly identical conditions. However, to achieve an equal CaO replacement rate a higher actual stone replacement rate was required since 1691 has only 81% of the CaO content of the UK stone. The efficiency vs cycle behaviour of 1691 was nearly identical to that of the UK stone. The lined out level of 1691 was also 72% sulphur removal.

These results are plotted in Figure 18 and listed in Tables C-1 and C-3, runs 54 - 72. It was not possible to use the same replacement rate in the initial cycles with limestone 1690. Preliminary tests showed lime losses in the first few cycles to be much higher than 0.75 wt CaO/wt.S Therefore the cyclic test was started at 1.5 wt CaO/wt S. This rate was maintained for ten cycles (Figure 19) by which point the efficiency had lined out at 80%. The high rate of solid losses encountered in the first cycle declined in subsequent cycles permitting a reduction to 0.75 wt CaO/wt.S replacement rate in the final ten cycles. The sulphur absorption efficiency then declined to about 64%. Data are listed in Tables C-1 and C -3, runs 78 - 95.

CYCLE TESTS (U.S. LIMESTONE 1691)

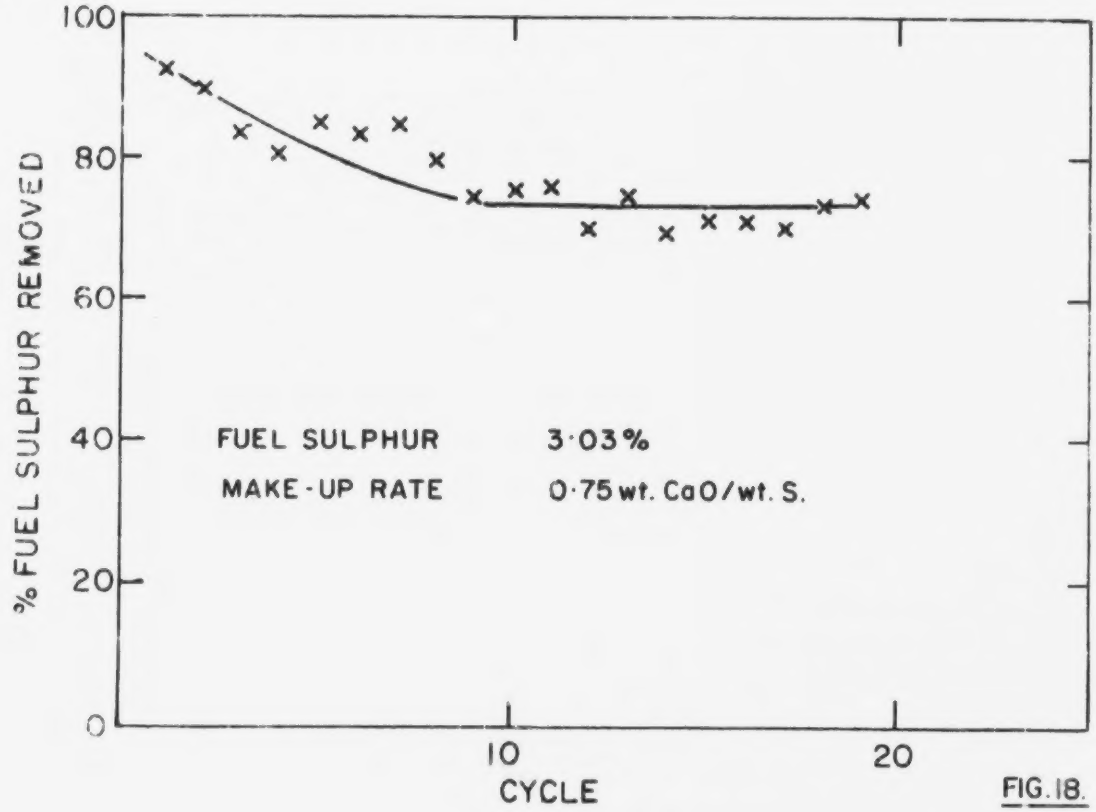


FIG. 18.

CYCLE TESTS (U.S. LIMESTONE 1690)

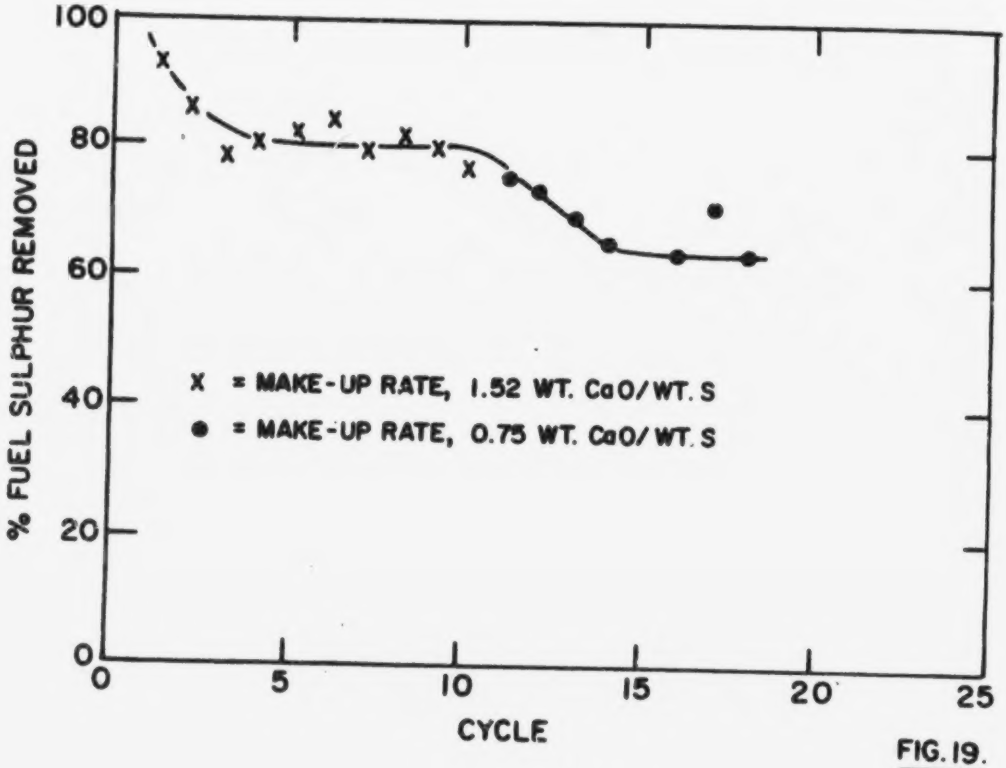


FIG. 19.

Limestone 1690 gave good sulphur removal efficiency in its first cycle, but in other respects it was inferior to 1691. Its low calcium content (60% of limestone 1691) gave it a low sulphur capacity on a total stone basis. It underwent severe attrition during calcination and in its initial gasification cycle. But even more serious, it tended to sinter and agglomerate at operating conditions. Figure 20 shows some of the agglomerates. This tendency to sinter and agglomerate would present severe operating problems to a continuous CAFB unit. It was not observed with any stone other than 1690.

AGGLOMERATES FROM BATCH REACTOR AFTER
CYCLE TEST WITH BCR1690 STONE

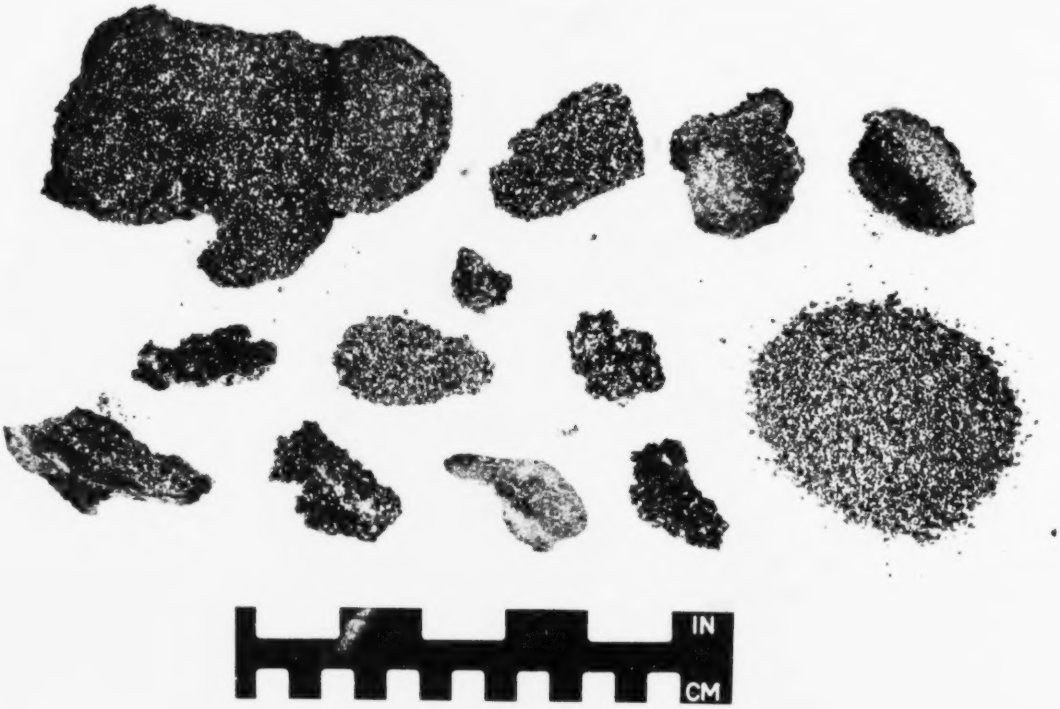


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FIG. 20.

Lime losses during the stone comparison tests were estimated by weighing the beds before and after the tests. Figure 21 shows lime losses per cycle with the UK stone and the two US stones. The total lime charge to the reactor was 8.2 kg.

HISTORY OF LIME LOSSES IN CYCLIC TESTS

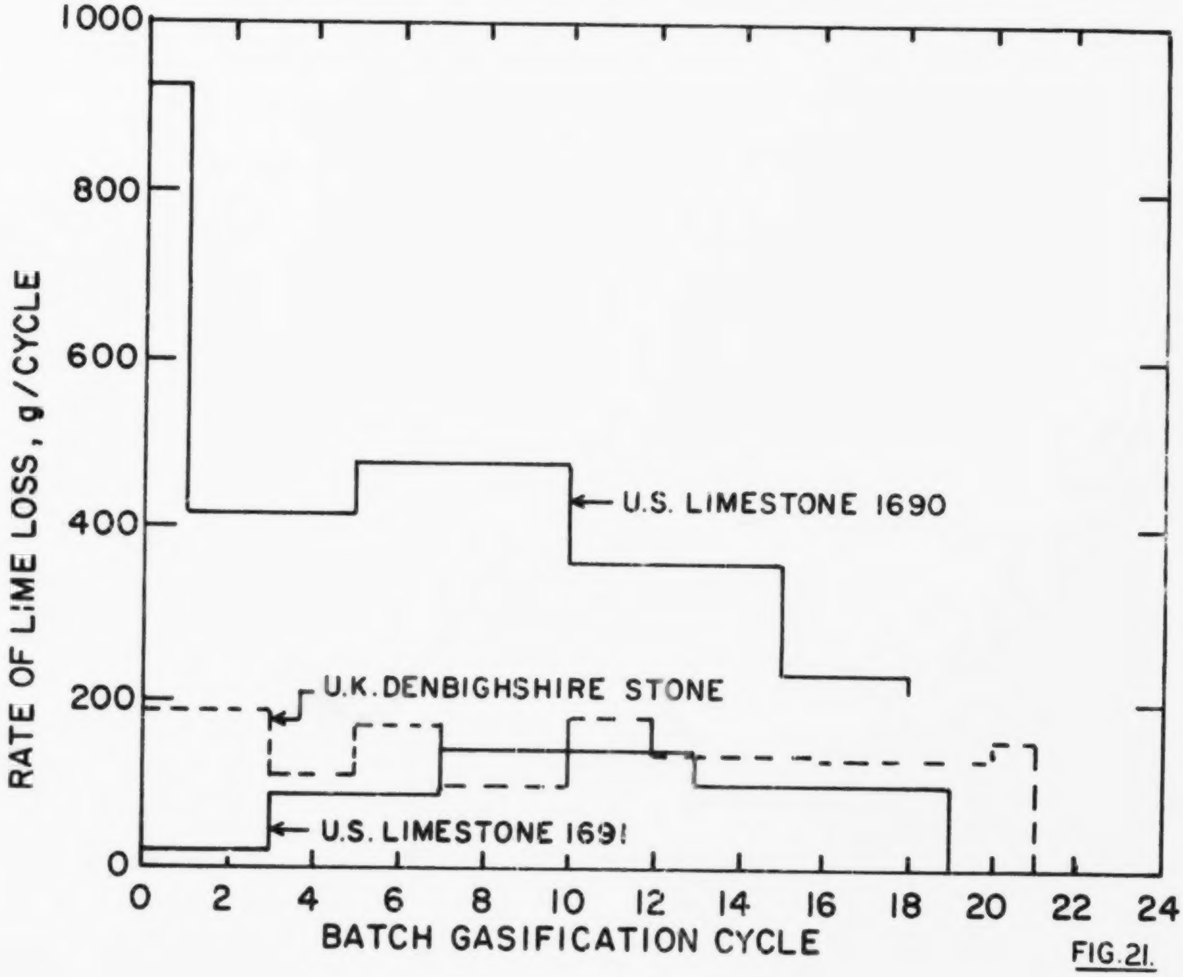


FIG. 21.

The UK stone settled down to a loss of about 140 g/cycle which was 1.7% of the bed or 1.4% on fuel fed. The US stone 1690 had a very high loss in the first cycle, but lower losses thereafter. The loss rate declined still further when replacement rate was reduced. During the final cycles the losses averaged 244 g/cycle, which is 3% of the bed or 2.5 wt % on fuel feed rate of 165 g/min.

Losses from stone 1691 started low, increased to a maximum at about cycle 10 and then decreased again. The average rate of loss for cycles 4 through 19 was 112 g/cycle which was 1.1% of fuel feed rate.

On the basis of these cyclic tests, limestone 1691 was selected as the better of the two stones suggested by NAPCA. Except for requiring a higher absolute make-up rate because of its lower CaO content, it appears to be as good as the UK Denbighshire stone for CAFB operations. On the other hand, stone 1690 does not appear very suitable for the CAFB process. It would require an even higher replacement rate, and it could cause serious operating problems.

2. Sulphur Removal During Gasification

Provision of sulphur-free fuel is the primary purpose of the CAFB process. Consequently major attention has been focussed on the factors which determine the sulphur removal efficiency, abbreviated SRE. We have found that gasification temperature, air/fuel ratio, fluidisation gas velocity, bed depth, limestone particle size, lime replacement rate, and the quantity of sulphur to which the lime is exposed each cycle, all affect the SRE. A number of variables interact with each other and therefore must be considered together.

a. Bed Processes During Gasification

The following visualisation of the various processes which must take place in the fluidised bed is an aid in interpreting the experimental observations made in this study. The chemical processes which occur in the CAFB gasifier are a complicated mixture of cracking, coking, oxidation, gasification, and absorption. Oil enters the bed some distance above the air distributor in a region where there may be little or no oxygen. The oil contacts the hot lime particles, is partly vapourised, and undergoes cracking and coking. Most of the organic sulphur compounds probably crack to H_2S which reacts with CaO in the upper portion of the bed to form CaS. Lime particles with a coating of coke and heavy hydrocarbon eventually circulate to the lower portion of the bed near the air distributor where the coke is oxidised. This oxidation supplies heat to the system and removes the coke that acts as a barrier to further sulphur absorption. Under some conditions, there may be enough oxygen in the vicinity of the distributor, if temperature is high enough, to convert some of the calcium sulphide to sulphate or even to oxidise the CaS to CaO and SO_2. In the latter case, the SO_2 would be reabsorbed higher in the bed.

b. Fresh Bed Sulphur Absorption Curves

The early part of a batch gasification cycle with a bed of fresh lime usually gives an SRE close to 100%. The duration of this high efficiency period and the subsequent variation of SRE with extent of lime utilisation depends strongly on gasification conditions. In the 1-A batch reactor it was not possible to separate effects of air/fuel ratio from those of gasification temperature as the two were strongly cross-correlated. As there was no independent temperature control, the gasification temperature increased as air/fuel ratio increased and oxidised a larger fraction of the fuel.

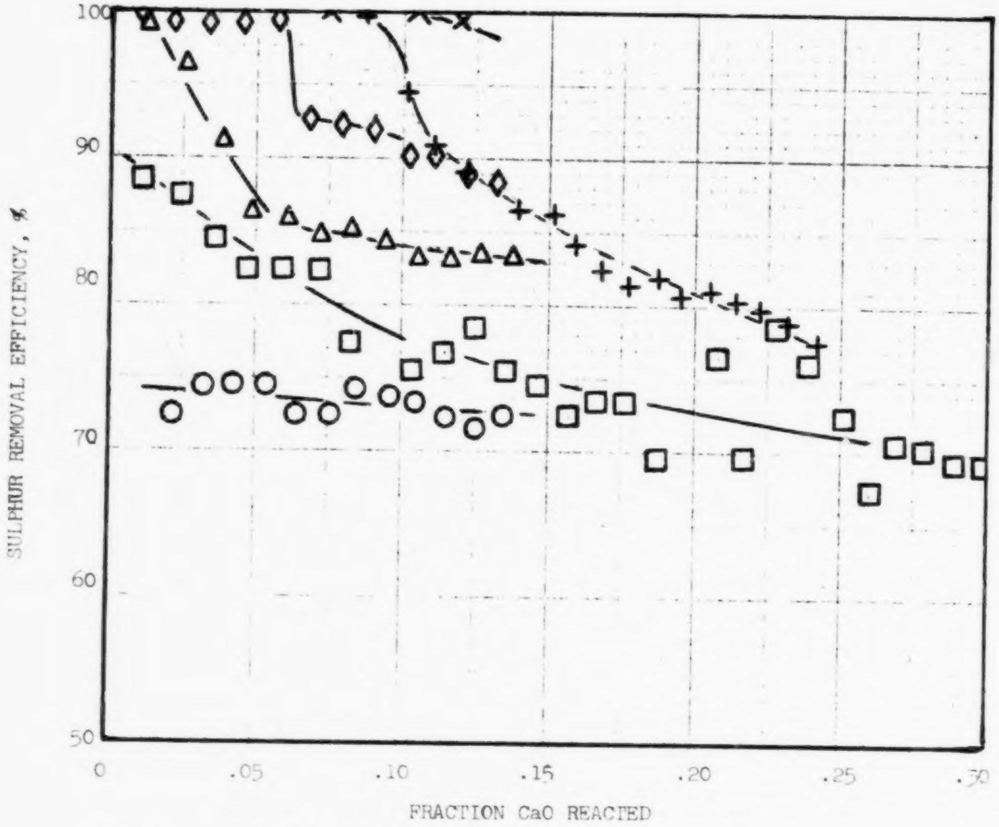
Figure 22 shows 1-A unit sulphur absorption curves at several air/fuel ratios and temperatures. The conditions and results are summarised in Table VI and reported in detail in Appendix Tables C-1 and C-3. At the lowest air/fuel ratios the SRE fell immediately to a low level and then began a further gradual decline. At intermediate conditions there was a period of high SRE followed by a sharp drop and a gradual decline. At the highest air/fuel ratio, the high SRE period persisted longer and the subsequent decline was more gradual. The low temperature-low air/fuel ratio conditions also were associated with high levels of carbon on the lime at the end of the cycle.

TABLE VI

Conditions for 1-A Unit Fresh Bed Tests Plotted in Figure 22

Run	Symbol	% Stoic Air	Temp °C	%C on Stone	Lime	Fuel
98	○	19.1	815	6.04	UK	UK
96	□	19.3	800	-	UK	UK
99	△	20.5	810	5.59	UK	UK
102	◇	22.8	860	1.71	UK	UK
103	+	24.8	865	0.15	US	US
121	x	26.7	850	.87	UK	UK

SULPHUR REMOVAL CURVES AT DIFFERENT AIR/FUEL RATIOS



Run	Symbol	% Stoich. Air	Fuel	Lime
98	○	19.1	U.K.	U.K.
96	□	19.3	"	"
99	△	20.5	"	"
102	◇	22.8	"	"
103	+	24.8	U.S.	U.S.
121	×	26.7	U.K.	U.K.

FIG.22.

Similar results, but at much higher levels of calcium utilisation, were obtained in the 4-A and 4-B reactors where better insulation permitted higher gasification temperatures. Figure 23 shows these curves. Again the low temperature-low air/fuel ratio test suffered rapid decline in SRE. But at 23.3% of stoichiometric air and 850°C, the lime maintained high SRE to over 20% calcium utilisation.

Conditions for these tests are listed below.

TABLE VII

Conditions for Sulphur Removal Tests in No. 4 Batch Reactors
6 ft/sec Air Velocity

Run	Symbol	% Stoic Air	Temp °C	%C on Stone	Lime	Fuel
24-4B	●	31.4	870	.04	US BCR 1691	Venezuelan (2.3%S)
27-4B	×	23.3	850	.38	"	"
6-4A	○	19.7	845	4.68	"	"
28-4B	△	14.8	780	11.88	"	"

RESULTS AT 6FT/SEC. IN NUMBER 4 UNITS
SULPHUR REMOVAL CURVES AT DIFFERENT AIR/FUEL RATIOS

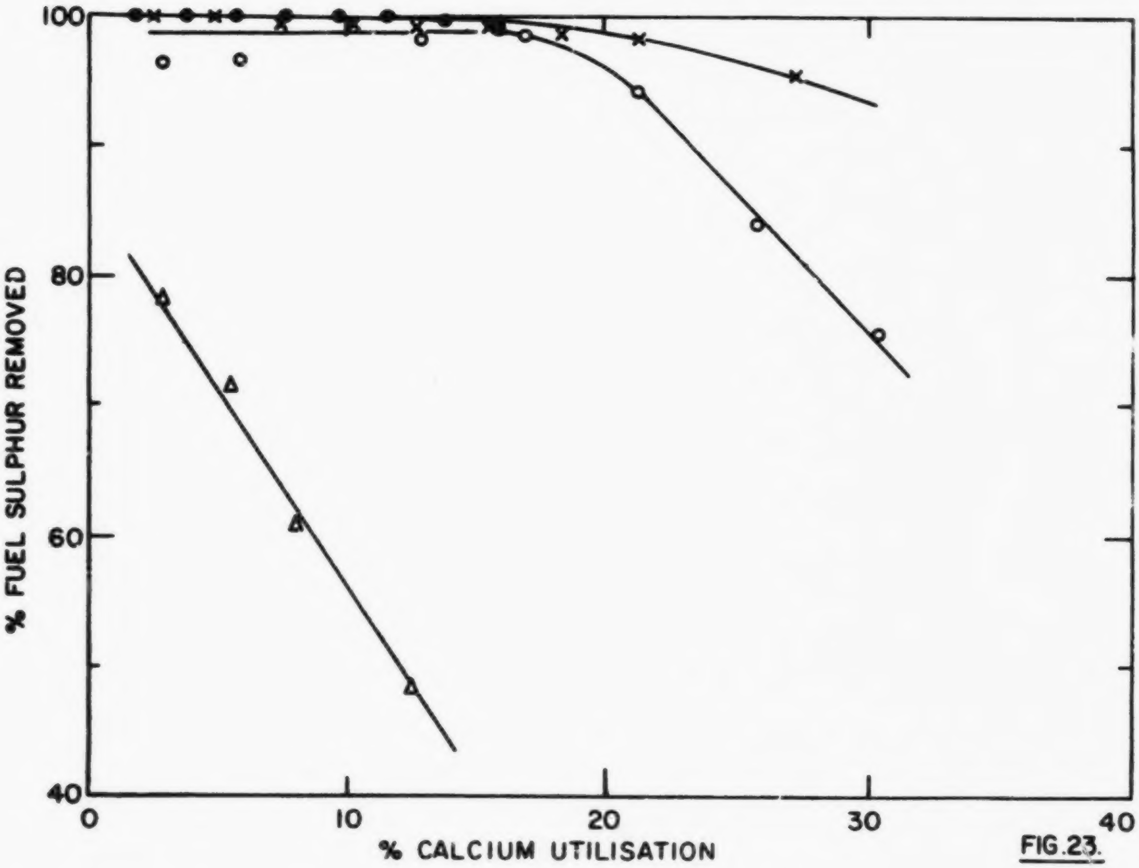


FIG.23.

These tests indicate that levels of calcium utilisation approaching 30% with maintenance of SRE over 90% are possible if proper gasification conditions are employed.

c. Temperature and Air/Fuel Ratio

Because the fuel handling capacity of a given size gasifier increases as air/fuel ratio decreases, efforts have been made to operate at lowest possible air/fuel ratios and to define the limitations that may exist. Use of cooling coils in the No. 4 units made possible the separation of temperature and air/fuel ratio effects. The fresh bed variable study with US fuel and limestone included tests over a range of air/fuel ratios at 800°C and at 840 - 870°C. Results in Figure 24 show that the higher temperature operations produced good SRE (measured at 5% S on lime) to below 20% stoichiometric air. However, at 800°C the SRE at 5% S declined at air/fuel ratios below 20% of stoichiometric, but was equal to the high temperature result at 25% of stoichiometric air.

INTERACTION BETWEEN AIR/FUEL RATIO AND BED TEMPERATURE

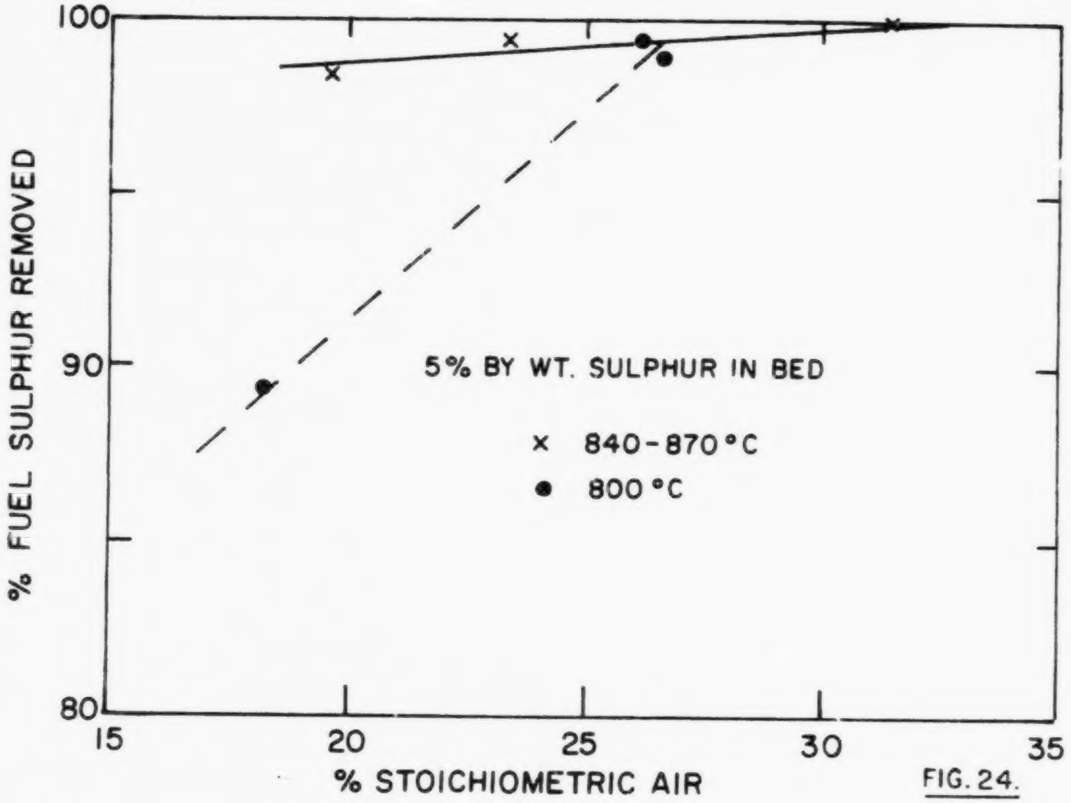


FIG. 24.

Table VIII shows that in these tests also, the level of carbon on stone increased sharply with lowered temperature and air/fuel ratio.

TABLE VIII

Stone Carbon Content in Temperature Comparison Tests

Run No.	Gasification Temp Average °C	Air/Fuel Ratio % of Stoichiometric	Carbon on Lime at Cycle End wt. %
24-4B	870	31.7	0.04
27-4B	850	23.3	0.38
6-4A	845	19.7	4.68
8-4A	800	26.1	1.59
26-4B	800	18.2	7.12

The sulphur absorption curves obtained in batch tests suggest that loss of sulphur removal efficiency can take two forms: (1) a very rapid loss bearing little relationship to the amount of sulphur reacted with the lime; and (2) a slow decline as the available lime is converted to CaS. The data indicate that the rapid loss in efficiency in some tests is due to building of a layer of coke or carbon on the reactive lime which acts as a barrier to sulphur absorption. Preferred CAFB operating conditions are those which avoid build-up of this carbon barrier.

Figure 25 shows carbon accumulation plotted as the ratio of net carbon deposition to oil fed versus air/fuel ratio, a form suggested by R. Newby.(7)

VARIATION OF CARBON DEPOSIT WITH AIR-FUEL RATIO

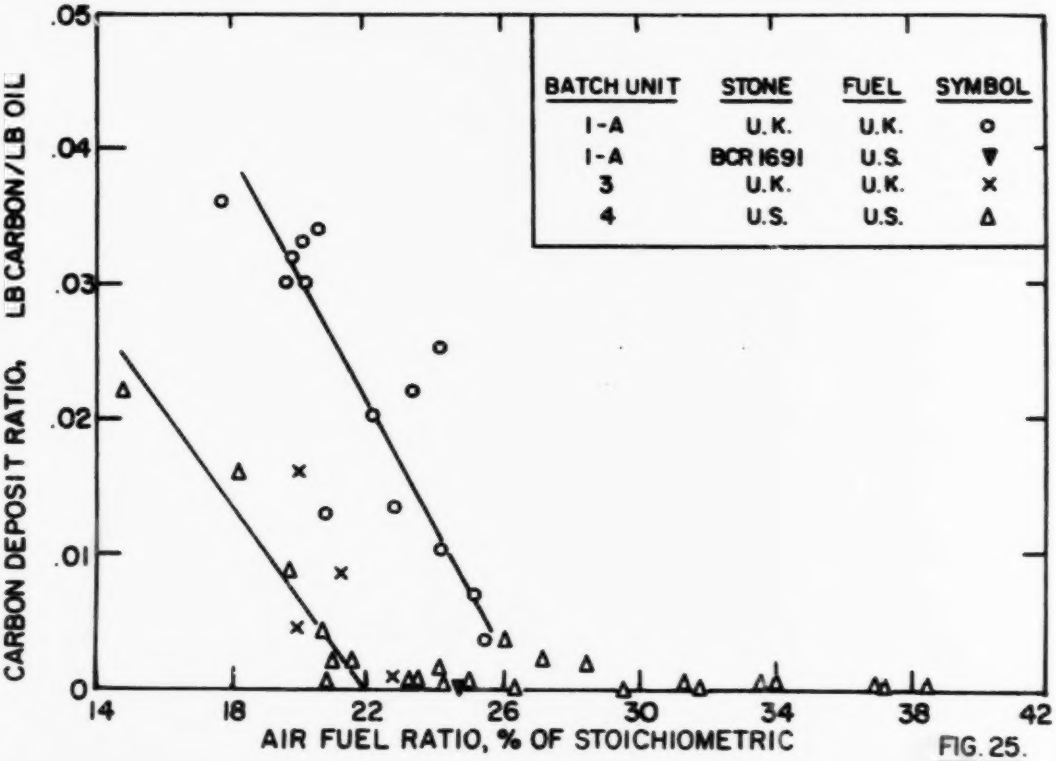


FIG. 25.

At high air/fuel ratios carbon accumulation is negligible, but below some critical level the amount of carbon laid down increases rapidly as air/fuel ratio decreases. Carbon levels observed in tests in the No. 4 units and the tall No. 3 unit were generally lower than those found in the No. 1-A unit, but the cause of this difference is not yet clear. Since temperature and air/fuel ratio were related in the batch units, similar curves are obtained if carbon is plotted versus temperature. The general behaviour of the data agrees with a mathematical model based on carbon oxidation kinetics.

This model predicts carbon-on-lime concentrations of the same magnitude as those actually observed and also predicts a steep rise in carbon concentration as temperature and/or air/fuel ratio decreases.

The model (described in Appendix K) assumes that coke is laid down from the fuel oil at a rate proportional to oil feed rate and Conradson Carbon concentration. It assumes that thin layer carbon burning kinetics apply; i.e. that carbon is burned off at a rate proportional to coke concentration in the lime, to oxygen concentration in the burning zone below the oil injector, and to an oxidation rate constant which increases exponentially with temperature. The effect of temperature on CO/CO_2 ratio in the product also increases carbon burning rate with temperature. The steady state carbon concentration is that which produces a coke burning rate equal to the coke deposition rate. Figure 26 compares measured concentrations of carbon-on-lime with curves predicted by the model for 15 and 30% of stoichiometric air at 4 ft/sec superficial velocity.

EFFECT OF TEMPERATURE ON CARBON CONTENT OF BATCH GASIFICATION LIME

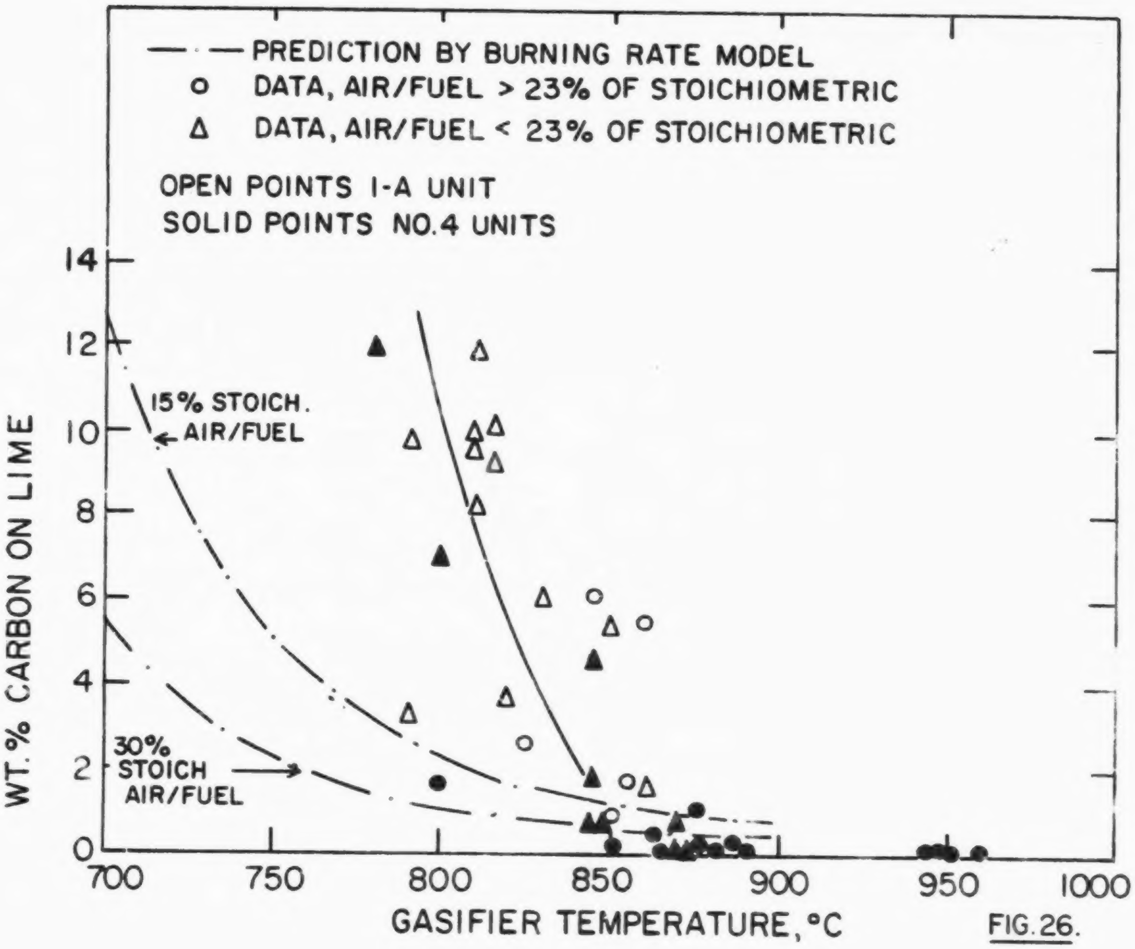


FIG.26.

Although the model predicts the correct directional trends, the data at low air/fuel ratio show much higher carbon levels than predicted below about 850°C. Failure of several assumptions of the model may cause this effect. For example, if a greater fraction of oil than Conradson Carbon deposits as carbon on lime in the upper part of the bed, there may be an oxygen limitation in the carbon burning region. In this case, the effect of temperature on CO/CO_2 ratio may be very important in affecting efficiency of oxygen utilisation. Also, once an appreciable fraction of the lime surface becomes carbon coated, thin layer carbon burning kinetics no longer apply, and burning rate ceases to increase in proportion with carbon concentration. When the lime is completely covered, burning rate will be independent of carbon concentrations. Under these unfavourable conditions, carbon content of the lime does not reach an equilibrium level but increases throughout the batch cycle.

Evidence for this type of behaviour was obtained in 1-A unit batch runs in which carbon content was measured on lime samples drawn during a long gasification cycle. Figure 27 shows these data. In the tests at temperatures of 850°C and lower, the carbon content increased throughout the cycle. At 860°C and 23.3% of stoichiometric air, the carbon content also increased. But at 860°C and 29.3% of stoichiometric, carbon level remained insignificant. Increasing temperature to 865°C permitted operation at 24.8% of stoichiometric air without increase of carbon content.

CHANGE IN CARBON CONTENT OF LIME DURING BATCH GASIFICATION CYCLE 1-A UNIT

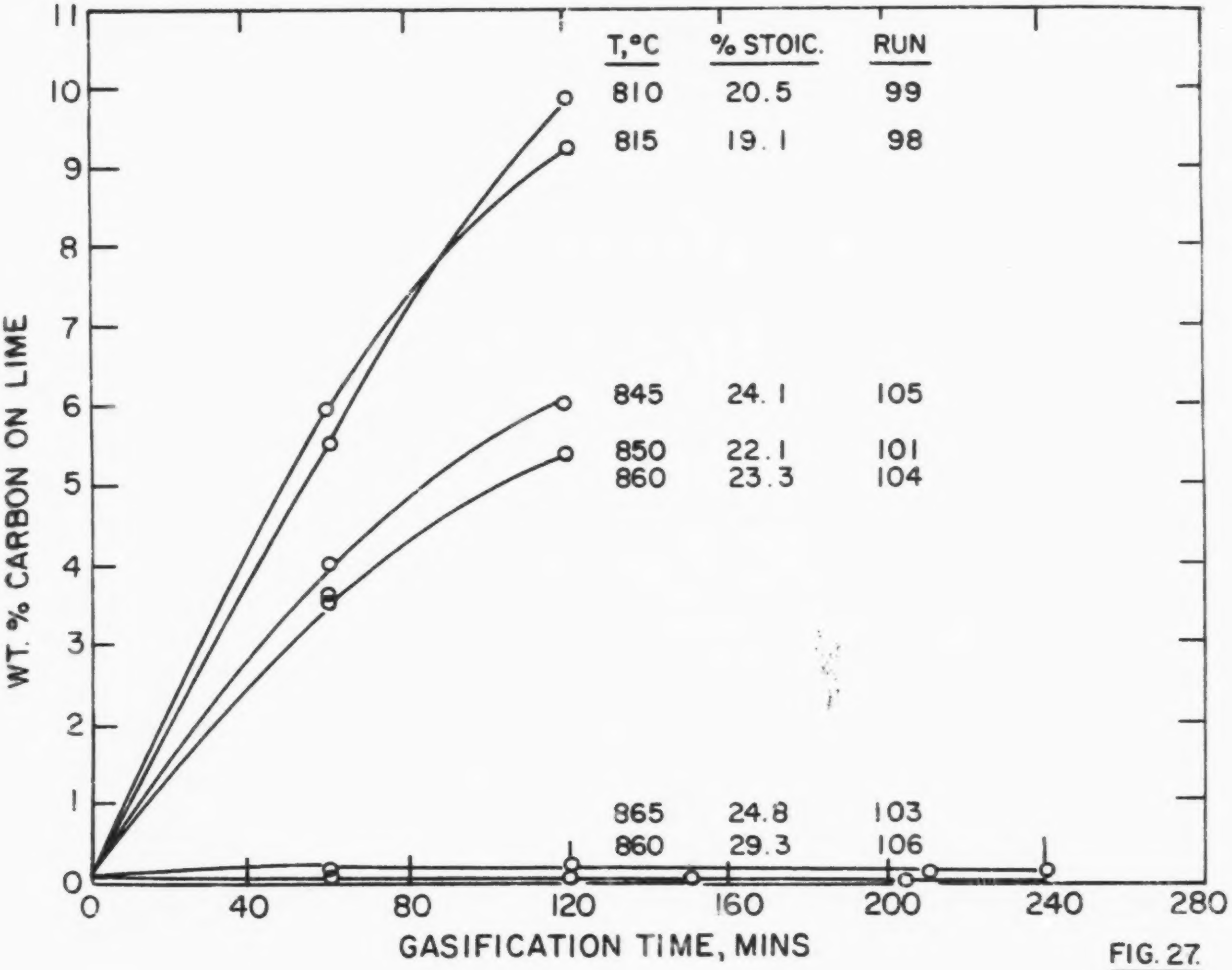


FIG. 27.

The scatter of data in Figures 25 and 26 and the large change in behaviour over a narrow range of conditions shown in Figure 27 indicate that the temperature range below about 870° and the air/fuel ratio range below 25% of stoichiometric is somewhat unstable. Results in Nos. 3 and 4 units and particularly, experience in the continuous pilot plant demonstrate that successful sulphur removal is possible at air/fuel ratios below 20% of stoichiometric provided that temperature is high enough. Figure 28 shows sulphur removal for a 15 hour period during the first test run in August. During this period air/fuel ratio averaged 17% of stoichiometric and SRE averaged about 95%. Gasifier temperature varied between 830 and 880°C.

LOG OF CAFB PILOT PLANT CONDITIONS AT LOW AIR/FUEL RATIO

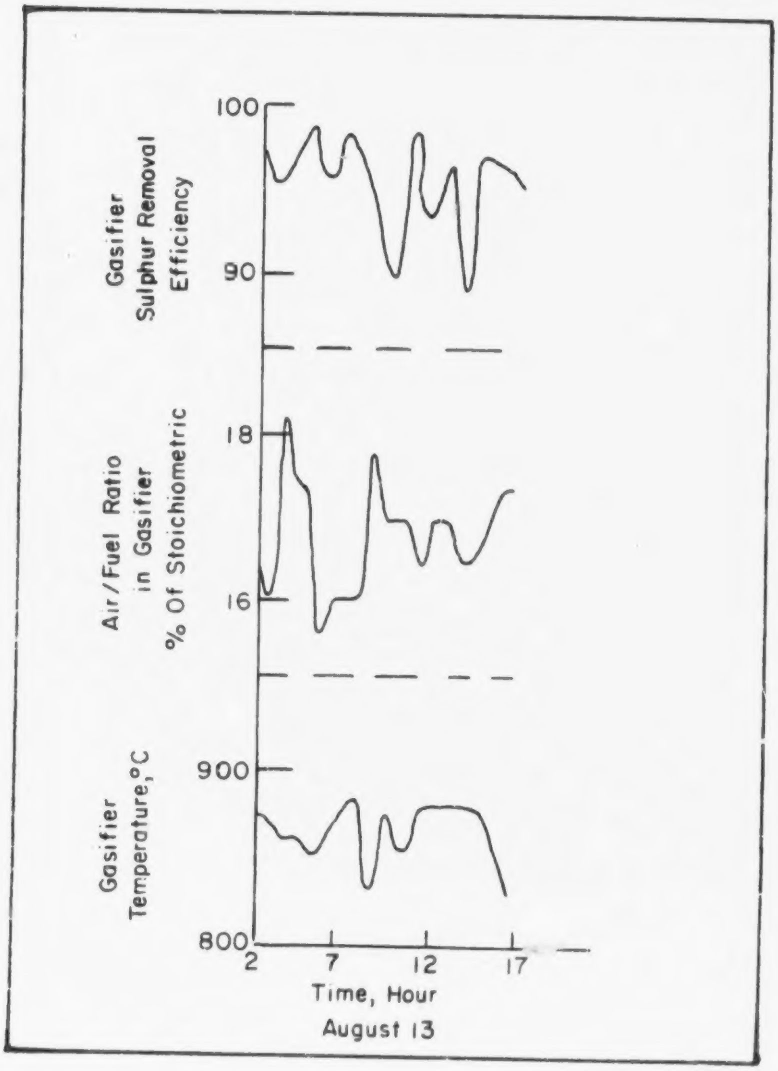


FIG. 28.

Carbon concentrations were not measured on gasifier solids in August, but data from the third run in November-December show that carbon concentrations on these solids averaged 0.1% with a measured high of 0.45%. These data appear in Appendix Table J-VII.

Because of its low surface area to volume ratio, the continuous pilot plant loses a lower fraction of its heat and can sustain high gasification temperature at much lower air/fuel ratios than the batch units. It appears that operation quite close to the adiabatic limit of about 15% of stoichiometric air will be possible without excessive carbon formation and loss of sulphur removal capacity.

The gasifier should be operated at 870°C or higher to provide a safe margin away from excess carbon laydown. In the batch unit variable study, conditions were selected to avoid undue carbon formation except in runs where air/fuel ratio and temperature were under study.

d. Upper Temperature Limitation

Although the equilibrium concentration of H_2S and other reduced sulphur compounds is very low at temperatures well over 1100°C, (Appendix A), experience has shown that sulphur removal efficiency decreases as gasifier temperature exceeds 900°C. Figure 29 shows this basic temperature effect on SRE.

BASIC EFFECT OF BED TEMPERATURE

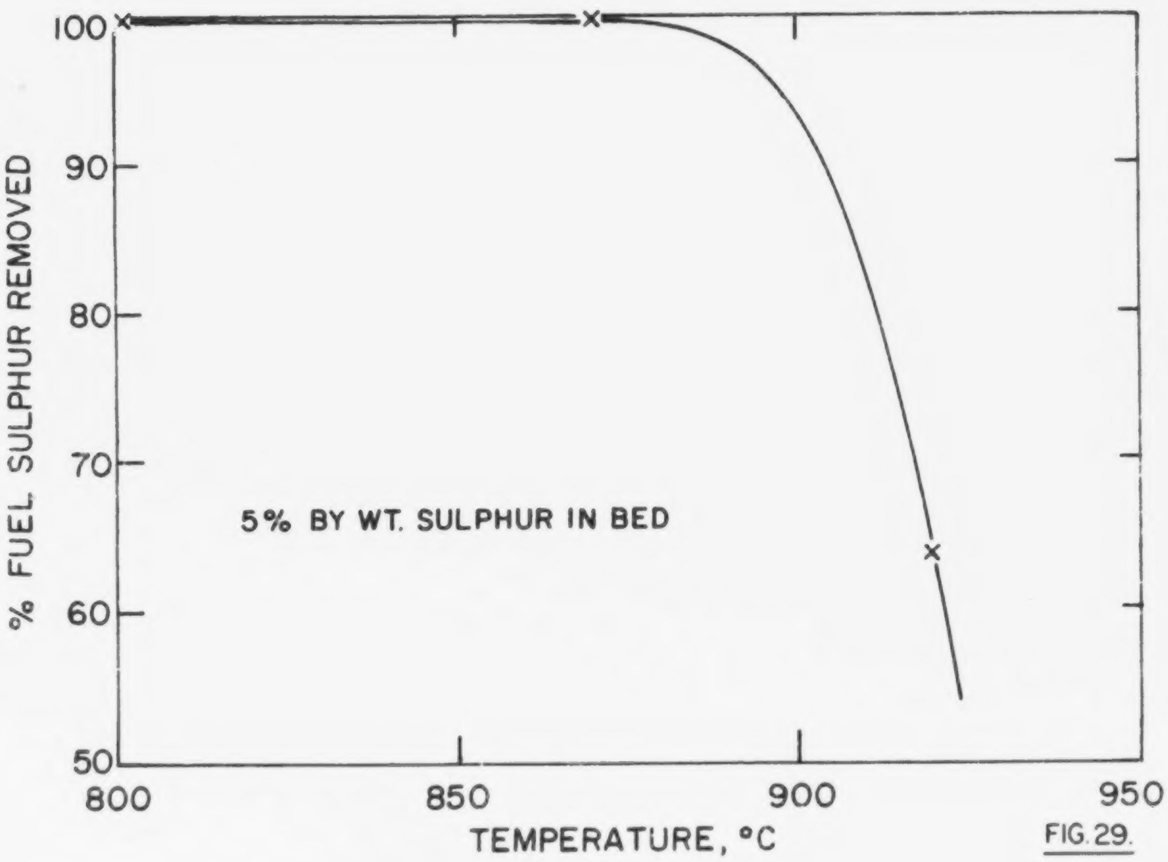


FIG. 29.

Batch test sulphur removal efficiency at 5 wt. % S on lime falls rapidly as temperature rises above 900°C. The explanation for this behaviour is thought to lie in the presence of an oxidising zone close to the distributor where CaS and the incoming air can react to form $CaSO_4$ and CaO + SO_2. The amount of SO_2 released increases with increasing temperature. It is likely that at temperatures above 900°C the quantity of SC_2 released is too large to be completely reabsorbed higher in the bed.

The optimum gasification temperature range is therefore 870 to 900°C to avoid excess carbon formation at the low temperature and local release of SO_2 at the high temperature.

e. Gas Velocity and Bed Depth in Fresh Bed Tests

In this work, the practice has been to discuss gas velocity in terms of superficial air rate - that is, the velocity which the air supplied to the reactor would have at operating temperature and pressure in an empty bed. The total gas velocity, of course, is different due to the presence of cracked oil products and the bed material itself. Bed depths also are normally expressed as static bed depths.

With other factors constant, and at an operating temperature of about 870°C, the basic effect of increasing air velocity is to decrease sulphur removal efficiency as Figure 30 shows.

BASIC EFFECT OF SUPERFICIAL GAS VELOCITY

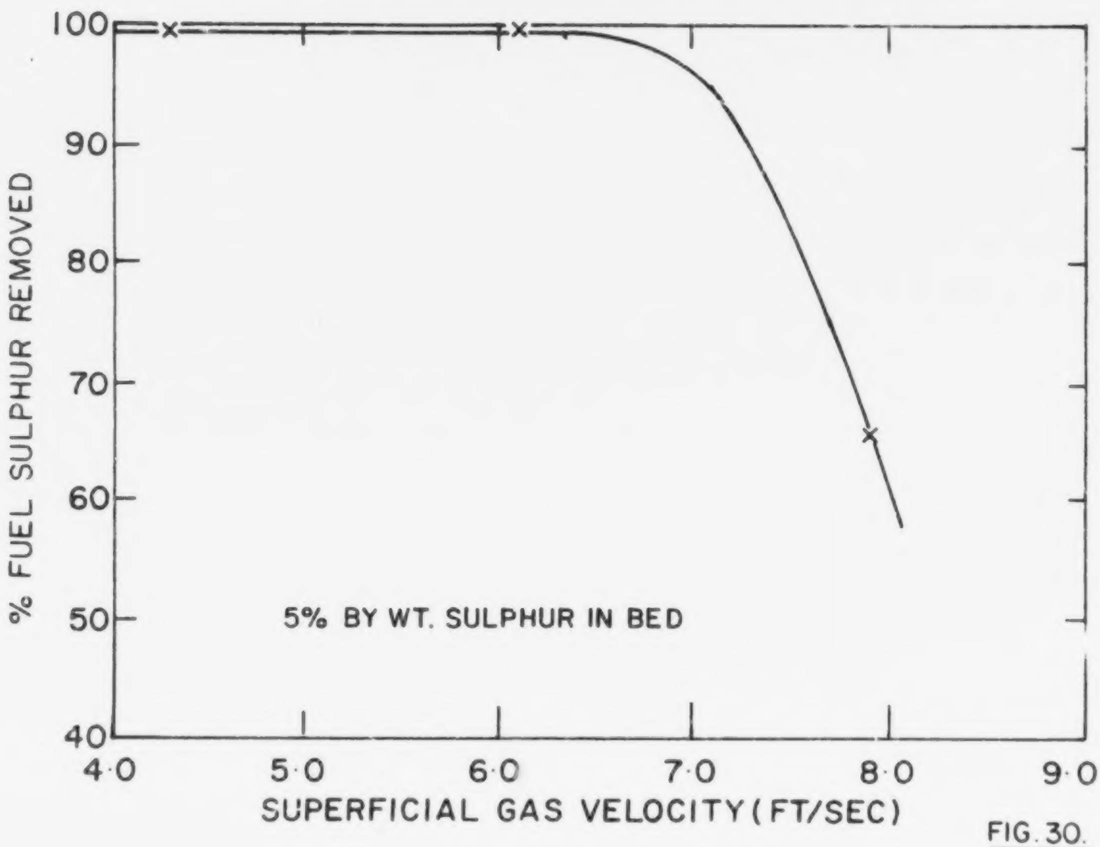


FIG. 30.

These data are from fresh bed tests with US fuel and limestone.

The SRE at 5% S in the lime remained close to 100% over the velocity range of 4.3 to 6.1 ft/sec but decreased to 66% at 7.9 ft/sec.

Changing bed depth at 6 ft/sec produces a similar effect shown in Figure 31.

BASIC EFFECT OF BED DEPTH

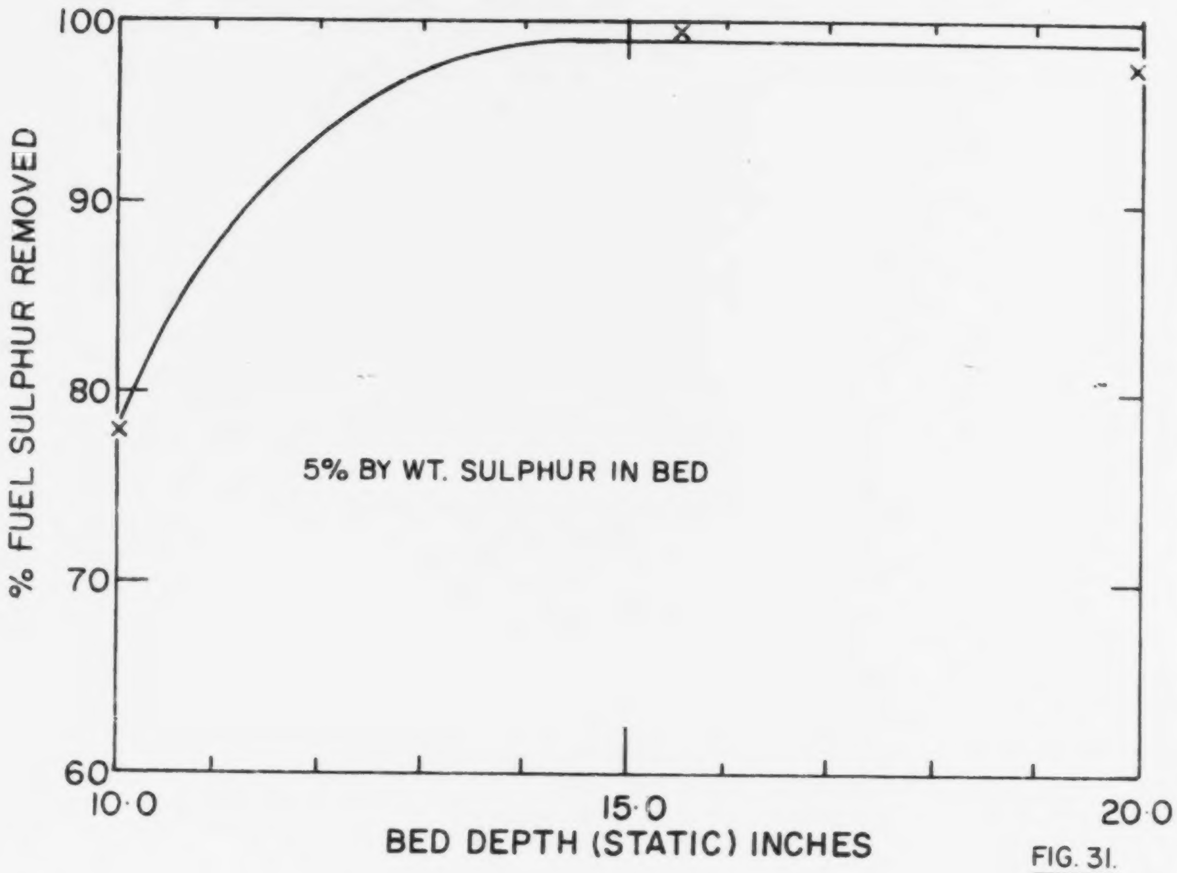


FIG. 31.

The SRE remained near 100% for static bed depths of 20 inches and 15.5 inches but declined to 78% at 10 inches depth. At 8 ft/sec the SRE was only 40% with a 10 inch bed but continued to improve with depth and reached nearly 100% with a 20 inch bed.

These gas velocity and bed depth effects can be correlated by consideration of the superficial gas residence time in the bed. Use of the residence time above the fuel injector gives a better correlation than total residence time as would be expected if there is little back-mixing of the gasified fuel downward from the injector. Figure 32 is a curve of SRE versus superficial gas residence time above

RELATIONSHIP BETWEEN SUPERFICIAL RESIDENCE TIME AND FUEL SULPHUR REMOVAL AT 5% BY WT. SULPHUR IN BED (FRESH BED TESTS)

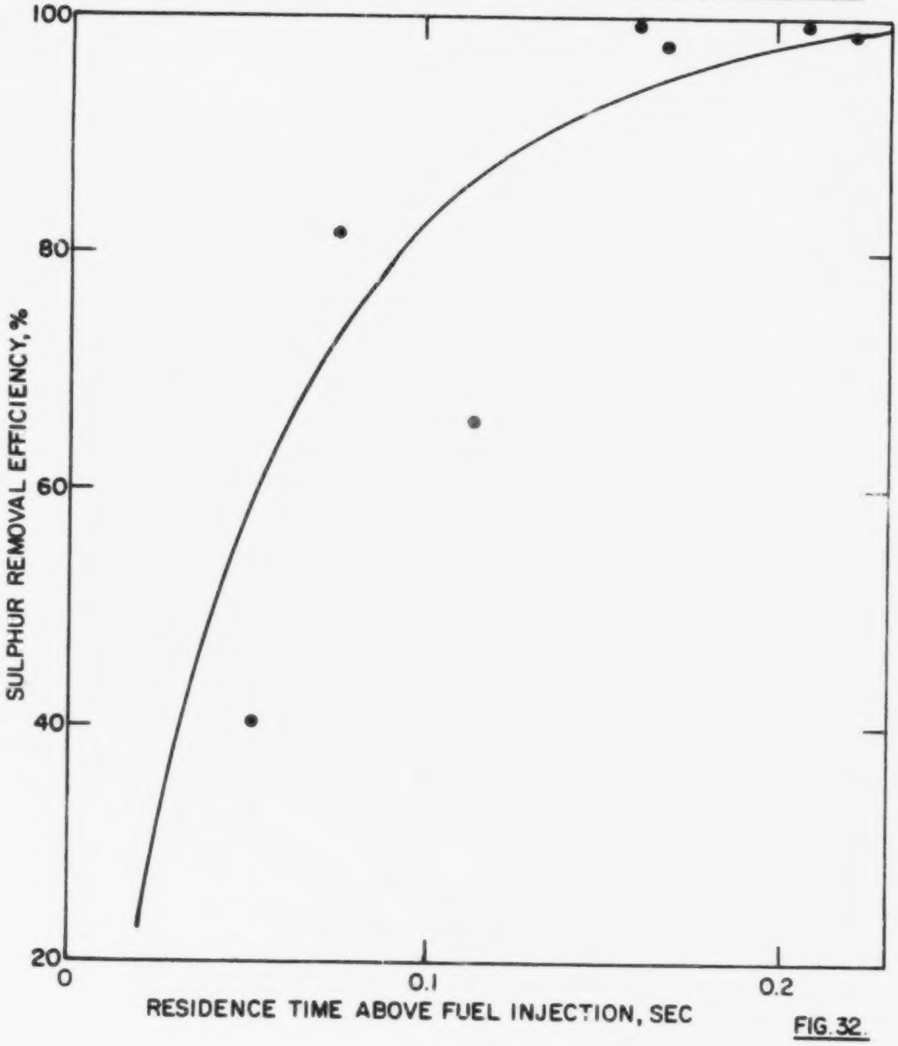


FIG. 32.

the oil injector. A first order reaction curve using a pseudo residence time approximates the behaviour of these data with a correlation coefficient of 0.88.

The first order rate constant under these conditions (870°C, 5% S in lime, starting from a fresh lime bed of 300 to 3175 microns) was 33.9 $\sec^{-1}$, and the pseudo time value was (t - 0.051) seconds where t is the superficial gas residence time above the fuel injector. The value of 0.051 subtracted from t is empirical, but may be due to the time required for fuel vapourisation and cracking before adsorption begins. Thus, the equation for sulphur removal efficiency in the fresh bed test at about 870°C is:-

$$SRE = 100 \left[1 - EXP\,(33.9\,(t - .051))\right] \qquad (1)$$

A reversal of velocity effect was observed in tests of the high gasification temperature of 950°C. At this temperature, sulphur removal is poor as discussed earlier. However, increasing velocity from 6 ft/sec to 8 ft/sec actually increased efficiency from below 20% to around 45%. This effect is not well understood, but possibly is due to increased bed circulation rate which minimised local temperature rise in the zone near the distributor and thus reduced re-release of SO_2 by local CaS oxidation.

Increasing bed depth at 950°C had the expected effect of improving sulphur removal efficiency. The test at 20 inches static bed depth and 6 ft/sec superficial air velocity gave 82% SRE at 950°C compared with values below 20% obtained with 15.5 and 10 inch beds.

f. Lime Particle Size

Early CAFB work had been done with narrow cut limestone size fractions of 600 - 1200 microns and 800 - 1400 microns. Yield of these narrow cuts is low in crushed stone, and for commercial economy a wider cut stone is preferred.

A $^1/_8$-inch to dust size fraction was considered from the standpoint of cost, but at velocities in the gasifier the low size end of this fraction would pass straight through the bed. Therefore a 300 micron cut point was selected for the lower particle size. Three particle size ranges were compared in batch unit fresh bed tests; 300-3175 microns, 600-1400 microns, and 1200-3175 microns. (The upper size of $^1/_8$-inch is approximate). A fourth size range, 600- 3175microns, has been used in some of the continuous pilot plant operations because of difficulty in screening large quantities of stone to 300 microns.

The results of fresh bed batch comparisons of the three selected stone sizes appears in Figure 33. The 300-3175 micron material maintained a high sulphur removal efficiency to nearly 30% calcium utilisation. The 600-1400 micron lime also gave reasonable performance but SRE with the 1200-3175 micron lime declined rapidly.

As Newby(7) observed, at low lime utilisation the relative performance of the three stones varied inversely as the square of their number average particle diameters. Simple reduction of particle external surface area with increasing particle diameter cannot account for the magnitude of efficiency loss observed. Some additional factor, such as diffusional resistance, must be included to explain the observed behaviour.

BASIC EFFECT OF PARTICLE SIZE RANGE

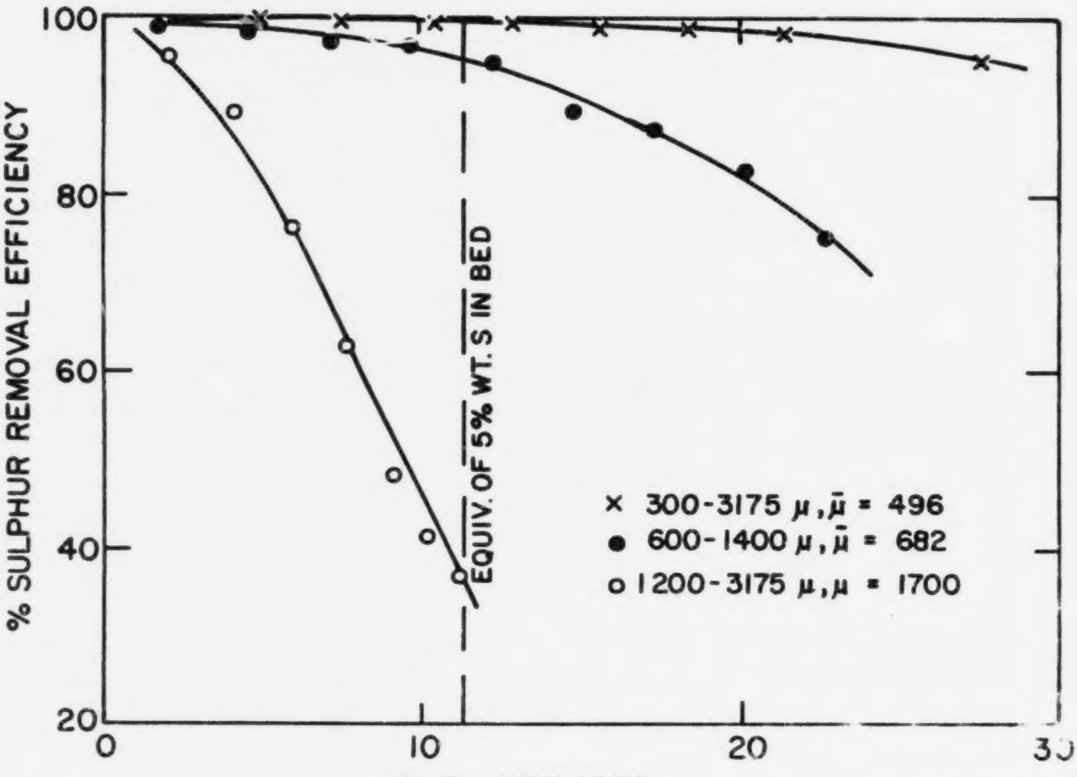


FIG. 33.

VARIATION OF SULPHUR REMOVAL EFFICIENCY WITH FUNCTION OF LIME UTILISATION

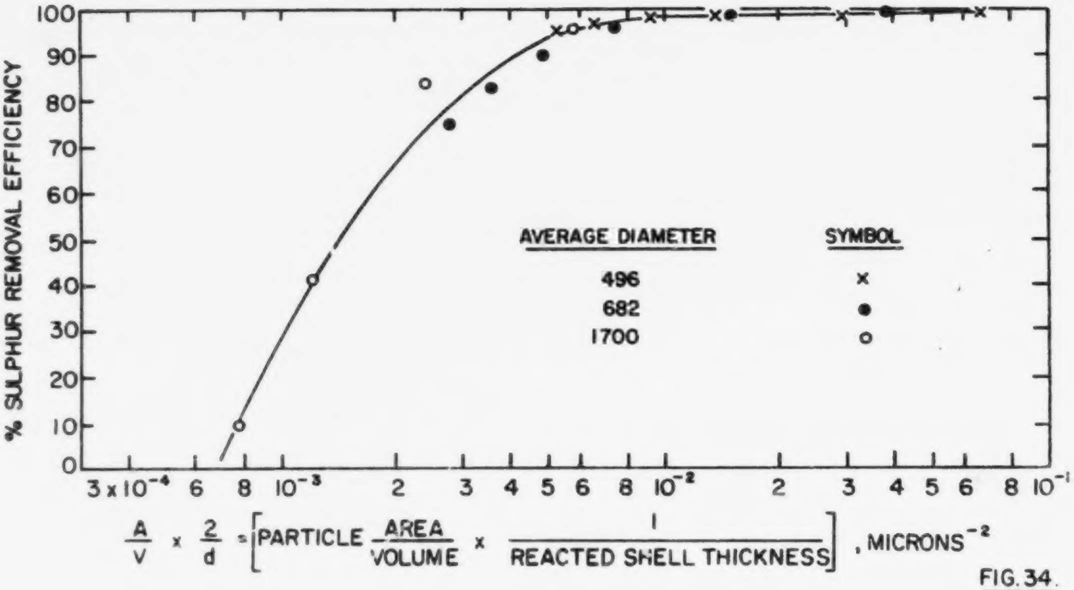


FIG. 34.

Figure 34 correlates test data from the three stones using the assumption that reacting lime particles are spheres with a shrinking core of unreacted material. The reaction rate at any extent of reaction is assumed to vary with surface to volume ratio divided by shell thickness. That is, SRE is a function of $\frac{A}{V} \times \frac{2}{d}$

where A = total particle external area
V = total particle volume
$D-d$ = diameter of unreacted core.

The fraction, U, of the sphere represented by the reacted shell is related to particle diameter, D, and shell thickness, $\frac{d}{2}$, by geometry.

$$U = 3\left(\frac{d}{D}\right) - 3\left(\frac{d}{D}\right)^2 + \left(\frac{d}{D}\right)^3 \qquad (2)$$

and

$$\frac{A}{V} = \frac{6}{D} \qquad (3)$$

The points in Figure 34 were determined by:

1. finding U at a given SRE and D from Figure 33
2. finding d from equation 2 at U and D
3. finding A/V from equation 3
4. calculating $\frac{A}{V}\frac{2}{d}$.

The relationship illustrated by Figures 33 and 34 shows that lime particle size can be traded for extent of lime utilisation to obtain a desired level of sulphur removal efficiency.

This relationship,of course, applies to fresh bed tests at the base line experimental conditions of 6 ft/sec superficial gas velocity and 15.5 inch static bed depth. As Table IX shows, increasing the bed depth improves fresh bed performance with the large size stone to an appreciable extent.

TABLE IX

Interaction between Bed Depth and Particle Size Range

	Static Bed Depth (inches)		
	10	15.5	20
% Fuel Sulphur Removed at 5% by wt. S in bed (300-3175 μ lime)	78.0	99.3	97.6
% Fuel Sulphur Removed at 5% by wt. S in bed (1200-3175 μ lime)	34.9	37.1	74.5

When adding fresh limestone under continuous operating conditions, the effect of particle size is complicated by lime losses. Cyclic test results on the effect of particle size differed from the fresh bed tests in that the finest size lime make-up did not give best results. Table X shows this comparison.

TABLE X

Effect of Limestone Particle Size in Cyclic Test

Test No.	Replacement Limestone Particle Size Range	Replacement Rate wt. CaO/wt. S	Lined Out SRE %	Sulphur Differential $\left(\frac{\text{wt. S fed}}{100 \times \text{wt. lime}}\right)$
T-3(Cycles 1 to 14)	300 - 3175	2.38	61	3.0
T-3 (Cycles 24 to 31)	600 - 1400	2.37	68	3.1

Limestone, BCR 1691;	Fuel, Venezuelan 2.3% S
Gasification Temp.	870°C
Air/Fuel	25% of Stoichiometric
Superficial Air Velocity	6 ft/sec
Static Bed Depth	15.5 inches
Gas Residence Time above fuel injector	0.15 sec

This effect was due to the higher relative losses of the finest particle size fraction during in-situ calcination of the stone. Separate tests revealed an average loss of 52.5% on calcination of 300-3175 micron stone compared with 42.7% of the 600-1400 micron stone. The weight loss due to CO_2 only is 37%. The effective make-up rate of the wide range stone, therefore, was less than that of the narrow range stone at equal addition rates. This effect did not appear in fresh bed tests where the reagent was added as precalcined lime.

In the third run of the continuous pilot plant, a limestone feed of approximately 600-3200 micron size was employed because of practical difficulties in screening the stone to the 300 micron cut point at high rates. Unless the stone is predried to a very low moisture content, a 300 micron screen blinds very quickly and gives large quantities of dust in the product.

g. Lime Replacement and Sulphur Differential

Two of the important variables in continuous CAFB operation cannot be studied in fresh bed batch tests but can be simulated in cyclic batch tests. These are lime replacement rate and sulphur differential. Replacement rate is the ratio of fresh limestone make-up to the quantity of fuel sulphur fed to the bed. Sulphur differential is the percent sulphur picked up on the lime between regenerations. In cyclic tests it is convenient to speak of projected sulphur differential (PSD) which is the increase in percent sulphur on lime which would result if sulphur removal efficiency were 100% throughout the cycle.

Projected sulphur differential is therefore:

$$\frac{(\%\ \text{S in fuel})(\text{fuel feed rate})(\text{gasification cycle length})}{(\text{wt. lime in gasifier})}$$

Lime replacement rate has been known as an important variable for some time. In the earlier CAFB batch studies it had been recognised that sulphur removal efficiency of a lime bed declined with repeated gasification-regeneration cycles but that the decline could be halted and a lined out SRE obtained if a portion of the lime is replaced after each cycle. The greater the replacement rate the greater is the lined out SRE. With the UK test fuel and stone at superficial gas velocity of about 4 ft/sec, a lime replacement of about 1.8 wt. S per wt. CaO maintained sulphur removal at nearly 100% efficiency, and replacement at 0.75 wt/wt gave 72% sulphur removal. An early test with U.S. stone BCR 1691(page 36) gave similar results, leading us to believe that the U.K. stone and BCR 1691 would give equal sulphur removal when replaced at equal CaO/S.

The initial cyclic tests in the current study with Venezuelan fuel and US BCR 1691 carried out at 6 ft/sec superficial air velocity and 300-3175 µ stone gave appreciably lower lined out SRE than those early UK fuel and stone tests. A check run (Test T-12) in one of the new batch units with the UK fuel and stone at conditions of the earlier tests produced nearly the same SRE as predicted (98% vs 92%). Thus the difference was not due to a change in equipment.

The reduction in gas residence time caused by increased velocity from 4 to 6 ft/sec, and the loss of small diameter stone during calcination discussed on page 57 cannot explain all of the efficiency loss observed. Additional factors must play a significant part.

But as yet, these factors have not been fully defined. Because of this unexpected lower efficiency at mild conditions, the cycle test variable study was extended to find conditions giving high levels of sulphur removal with the BCR 1691 stone without excessive lime replacement rates and low gas velocity.

Increasing bed depth at equal cycle length was found to improve efficiency out of proportion to the change in residence time. This increased benefit of a larger lime bed can be explained by the reduction in PSD which occurred. Lime cycled through a lesser change in sulphur concentration appears to retain greater absorption activity.

The effect of PSD was verified by reducing cycle length at constant bed depth. Table XI shows this comparison.

TABLE XI

Effect of Sulphur Differential (PSD)

Test No.	Lime Replacement Rate wt. CaO/wt. S	PSD %	Lined out SRE %
T-1 (Cycles 1 - 14)	2.43	3.0	61
T-9	2.55	1.9	86

Limestone BCR 1691; Fuel, Venezuelan 2.3% S.

Gasification Temp.	870°C
Air/Fuel	25% of Stoichiometric
Superficial Air Velocity	6 ft/sec
Static Bed Depth	15.5 inches
Gas Residence Time above fuel injector	0.15 sec

This large effect of sulphur differential implies that highest possible lime circulation rates between gasifier and regenerator should be used to improve sulphur removal. Heat balance considerations, of course, will determine the minimum practical sulphur differential.

h. Correlation of Fresh Bed and Cycle Test Results

A single fresh bed test is much quicker and easier to perform than a series of cyclic tests. It therefore would be very desirable to predict cyclic test and continuous unit performance on the basis of a fresh bed test. To explore this possibility, a function derived from the first order rate relationship used for fresh bed tests was modified to include effects of bed replacement rate and sulphur differential. The derivation appears in Appendix F. The final expression is:

$$\text{SRE} = 100 \left[1 - \frac{1}{10^{\frac{0.10\,k\,\theta\,m}{d}}} \right] \qquad (4)$$

where:

SRE = % sulphur removal efficiency

k = first order rate constant from fresh bed tests = 33.9 sec^{-1}

θ = (gas residence time above fuel injector - .051) sec.

m = Lime replacement rate, wt.CaO/wt. S

d = Projected sulphur differential, %.

Since the rate constant was determined for a particular set of conditions, this equation applies to the Venezuelan fuel - BCR 1691 limestone combination at temperatures in the range 840-900°C, air/fuel ratio of 20% to 30% of stoichiometric, and limestone of 300-3175 micron particle size range. This function fits the cyclic test data with a correlation coefficient of .85 but, as Figure 35 shows, usually predicts slightly lower values for SRE than observed.

CORRELATION BETWEEN EXPERIMENTAL AND CALCULATED LINED OUT EFFICIENCIES AND USING EQUATION. 4.

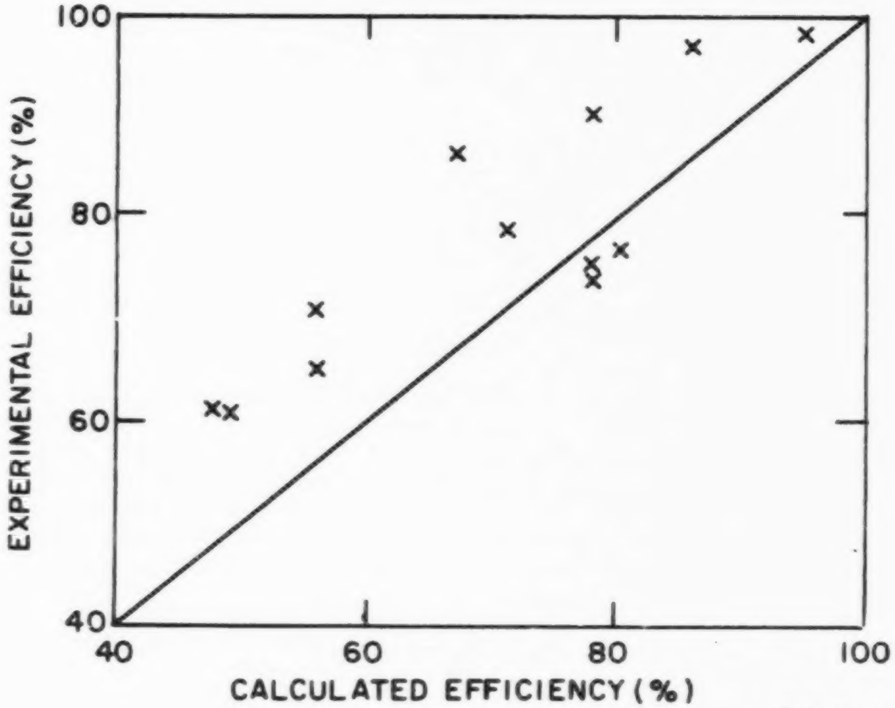


FIG. 35.

In order to obtain a more precise correlation of the actual cycle test data, a regression was performed using a function which, though similar to equation 4, permitted assignment of different weights to the separate variables.

$$\mathrm{SRE}\ \% = 100\left[1 - \frac{1}{\mathrm{EXP}\ \dfrac{w t^{x}\ m^{y}}{d^{z}}}\right] \qquad (5)$$

This function contains four adjustable coefficients or exponents whereas the constants in equation 4 were determined without reference to the cycle test data. In comparison with equation 4, the term $\underline{w}$ contains the $\underline{a}$ x $\underline{k}$ product while x, y, and z are weighting exponents for gas residence time, lime replacement rate, and potential sulphur differential.

Regression of the cyclic test data with equation 5 yielded the following result with a correlation coefficient of 0.94.

$$\mathrm{SRE} = 100\left[1 - \frac{1}{\mathrm{EXP}\ \dfrac{9.285\ t^{.6}\ m^{.95}}{d^{1.9}}}\right] \qquad (6)$$

The experimental results indicate, therefore, that potential sulphur differential has a more marked effect on lined out sulphur removal efficiency than suspected and that gas residence time has less.

Efficiencies calculated for a variety of values of the independent variables are listed in Tables KI through KIII. Figure 36 shows the effects of residence time and make-up rate at the reasonable potential sulphur differential of 2 wt. %.

The figure indicates a requirement for gas residence time greater than 0.3 sec to achieve greater than 90% sulphur removal with stoichiometric replacement of BCR 1691 limestone at 300-5175 μ particle size range.

Further work, especially with deeper beds, is necessary, however, to prove the reliability of equation 6 for extrapolation.

PREDICTED LINED OUT SULPHUR REMOVAL EFFICIENCIES (PSD 2·0 WT. %)
(BCR 1691) (300-3175μ)

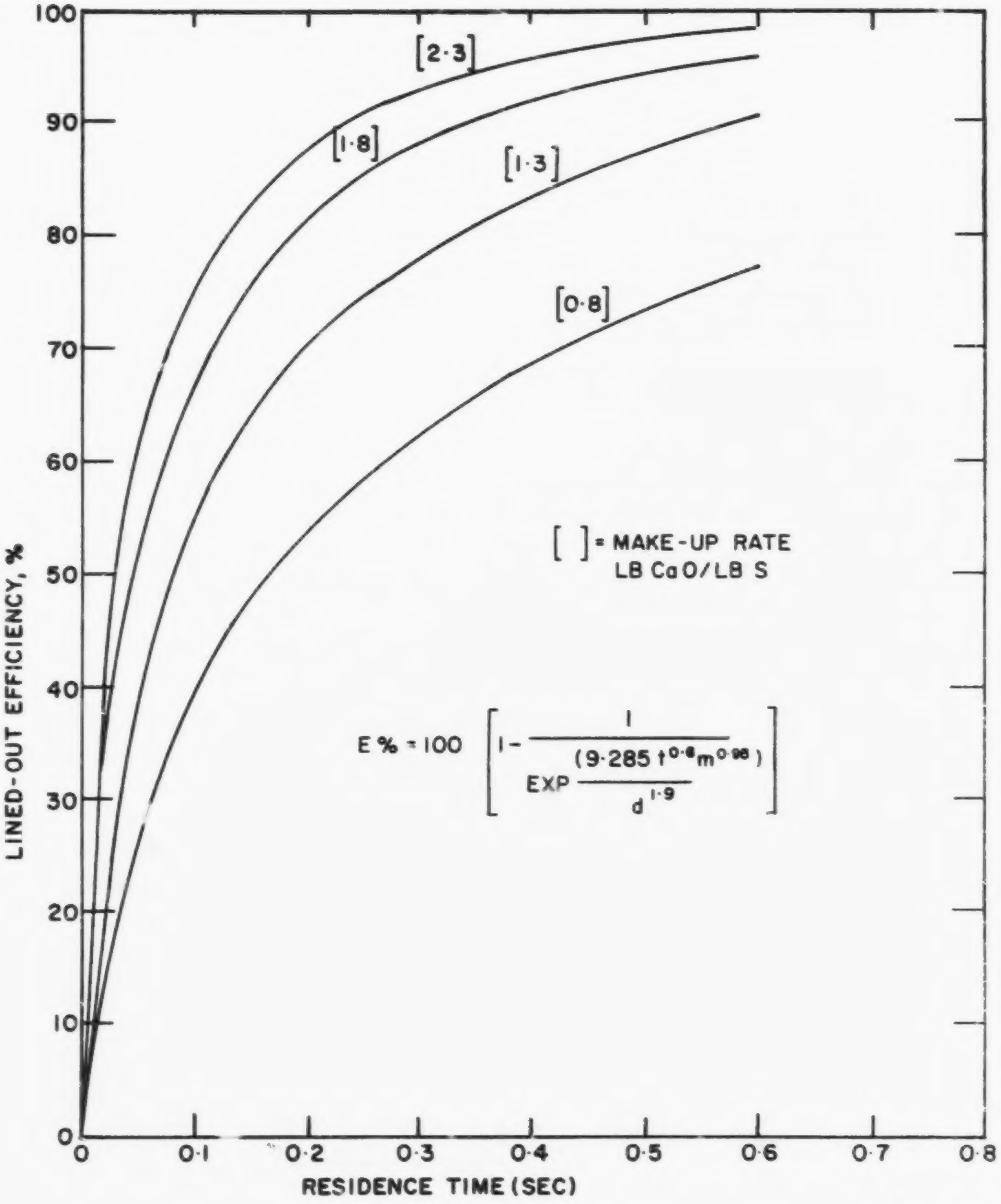


FIG. 36.

i. Continuous Pilot Plant Experience

Appendix J(1) contains results of the three runs of the continuous pilot plant. Figures J-1 and J-16 show the variation of sulphur removal efficiency with time during tests 1 and 3. In the continuous run period of the first test, SRE ranged from about 80 to 100% with an average of 90%. Lime replacement rate varied from about 1.1 to over 3 moles/mole S.

Sulphur removal from the fuel gas was essentially 100% throughout the second test in spite of beddepths as low as 11-12 inches w.g. However, lime replacement was quite high, ranging from about 3 to 4.5 moles per mole of sulphur, and a portion of the sulphur reacted with lime that passed through the boiler. In the final test run, sulphur removal covered the range from 75% to 100% with changes in bed depth appearing to be a major cause of the variations observed. There appears to be little difficulty in achieving SRE of 95 to 100% at stoichiometric lime replacement rate and gas velocity of 4 ft/sec with the Denbighshire stone used in these pilot plant tests.

3. Regeneration Behaviour

The function of the regeneration step is to convert sulphur to a concentrated SO_2 stream while restoring the lime to a form capable of active sulphur absorption. During the regeneration step the lime must be heated from gasification temperature of 850-900°C to regeneration temperature of 1000-1100°C. Energy for this heating comes from oxidation of carbon and sulphur on the stone. At the lower temperature range, this sulphur oxidation proceeds to form calcium sulphate.

The two principal reactions are:

$$CaS + 2\,O_2 \longrightarrow CaSO_4$$

$$CaS + {}^{3}/{}_{2}\,O_2 \longrightarrow CaO + SO_2$$

At high temperatures a third reaction comes into play:

$$CaS + 3\,CaSO_4 \longrightarrow 4\,CaO + 4\,SO_2$$

a. Fresh Bed Tests

A typical gas composition profile for batch regeneration is shown in Figure 37 (Fresh Bed Test 3).

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(1) Figures in Appendix J are located on pages 291 through 334.

EXIT GAS COMPOSITION DURING REGENERATION
(TEST 3)

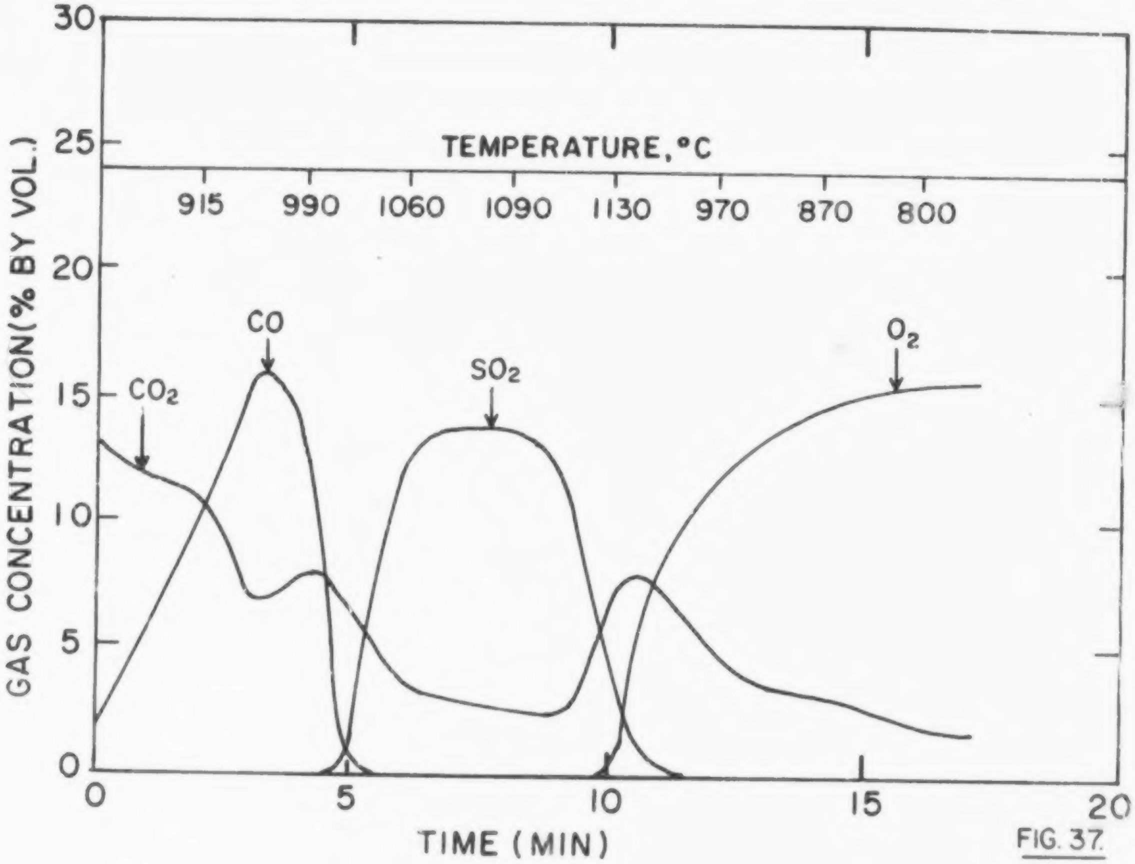

FIG. 37

There is an initial high carbon oxide evolution as carbon is burned off the lime, and the temperature rises. SO_2 then increases and reaches a maximum at temperatures between 1060°C and 1100°C. Then, as the reaction with CaS nears completion, oxygen begins to break through, and CO_2 increases as carbon deposited higher in the reactor is burned. When oxygen breaks through, the bed temperature starts to fall. At this point all the residual sulphur in the bed is present as calcium sulphate.

After each regeneration in the fresh bed test programme, a stone sample was analysed for sulphur content. These analyses revealed that the sulphur level varied between 0.35 and 3.14% by weight. The results of a correlation of sulphur contents with operating variables are summarised in Table XII.

TABLE XII

Operating Variable	Effect on Residual Sulphur Level
Gasification Temperature	Inverse
Gas Velocity	Direct
Air/Fuel Ratio	Direct
Particle Size Range	Direct
Bed Depth	None.

The operating variables are listed in descending order of importance.

The effects of gasification temperature, gas velocity, and air/fuel ratio during gasification all can be explained by their influence on the amount of exposure of sulphided lime to oxygen at temperatures below 1000°C. If gasification temperature is low, there must be more CaS oxidation to $CaSO_4$ before temperature is high enough to liberate SO_2. At low air/fuel ratio there is greater lay down of carbon, and oxidation of this carbon supplies heat to raise lime temperature with less requirement for sulphate formation. Increasing air velocity increases the average concentration of oxygen in the bed during the low temperature period and this increases conversion to sulphate. It is not understood, however, why particle size of the lime should influence the amount of sulphate formed during regeneration.

b. Cycle Test Regeneration

In cyclic tests, loss of absorption activity with repeated cycles did not reduce the ability of the stone to be regenerated. However, increasing the depth of the fluid bed did reduce the efficiency of regeneration. This result was caused by the greater heat loss from the deeper bed. This heat loss affected regeneration indirectly by forcing lower temperatures and higher air/fuel ratios during gasification, and it also produced a direct effect by removing heat during regeneration.

c. Continuous Regeneration

In a continuous regenerator the total bed operates at approximately steady state conditions. However, the individual lime particles must go through a cycle similar to that of a batch regeneration. The particle enters at gasifier temperature, is heated by carbon and sulphide oxidation, and finally releases sulphur as SO_2 before returning to the gasifier. Since air and gasifier lime both enter at the bottom and leave the top of the regenerator, there is a gradient of increasing temperature from bottom to top.

(1) Regeneration Selectivity

The regenerator temperature gradient reduces the selectivity of CaS oxidation to CaO + SO_2 by increasing the proportion of $CaSO_4$ formed. This behaviour is shown in Figure 38 where oxidation selectivity in the pilot plant regenerator is compared with calculated equilibrium curves. Selectivity was calculated from regenerator gas analysis. It is evident that the pilot plant regenerator required much higher operating temperatures than equilibrium to achieve a given selectivity. It also appears that somewhat better selectivity was obtained in Run 1 than in Run 3. The cause of this difference is not yet known.

COMPARISON OF CaS OXIDATION SELECTIVITY WITH EQUILIBRIUM CURVES

CAFB CONTINUOUS PILOT PLANT DATA

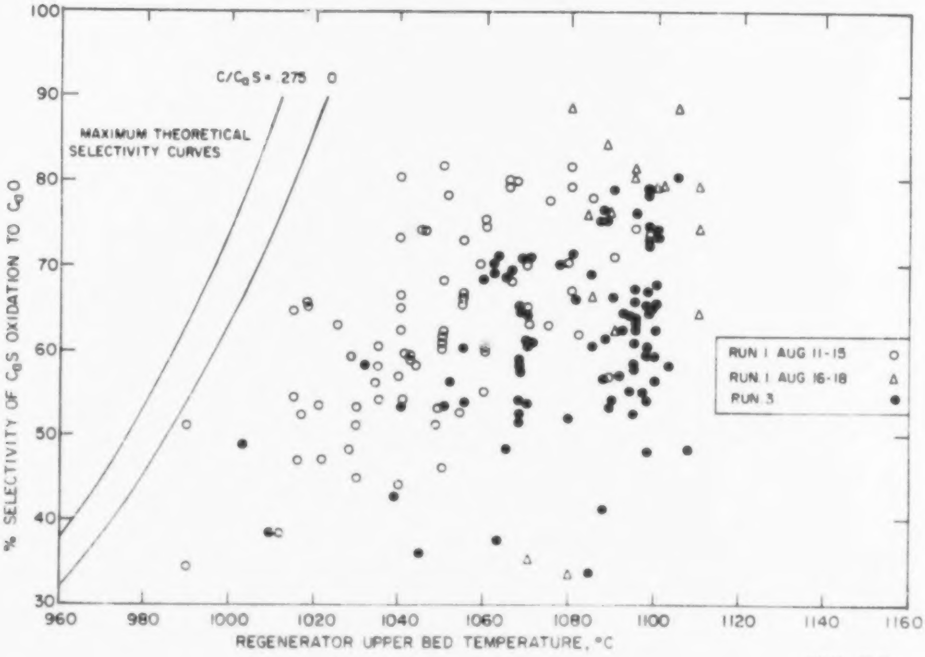

FIG. 38.

This reduction in regenerator selectivity has several undesirable effects. It reduces SO_2 concentration in the regenerator off gas thereby increasing cost of sulphur recovery. It raises the required regenerator temperature which probably accelerates lime deactivation. And finally it increases the amount of sulphur recycled to the gasifier which requires that the lime operates at a higher total sulphur loading.

(2) CaS Conversion

Not all of the calcium sulphide entering the regenerator is oxidised. Samples removed from the regenerator show appreciable quantities of unreacted CaS. It could be argued that since samples of solids were taken from the reactor itself and not from its outlet, they do not show the true extent of conversion. On the other hand, solids circulation rates, Figure J-30, estimated from sulphur concentrations of gasifier and regenerator samples together with the SO_2 production rate of the regenerator in some cases agree with, and in other cases are less than circulation rates estimated from regenerator heat balance. This comparison suggests that solids leaving the regenerator were not more completely reacted than indicated by the samples. Table XIII lists measured sulphur concentrations of gasifier and regenerator samples and the conversions and selectivities calculated from them.

TABLE XIII

Regenerator Performance Based on Sample Analysis
Continuous Pilot Plant Run 3

Time	Regenerator Solids		Gasifier Solids		Sulphur Differential %	% Conversion of CaS	% Selectivity to CaO
	Sulphide	Sulphate	Sulphide	Sulphate			
Day 1-1730	2.45	.95	7.52	.09	4.21	67.5	83
Day 2-1900	2.67	.57	5.26	.19	2.21	49.1	85
Day 3-0230	2.53	.64	6.00	.15	2.98	50	86
Day 3-0730	2.40	.65	4.95	.13	2.03	51.5	80
Day 5-0830	1.87	.84	3.72	.19	1.21	49.8	65
Day 7-1130	1.19	1.18	3.92	.23	1.78	70	65
Day 8-2230	.95	2.55	5.11	.24	1.85	81.5	44.5

Selectivities based on these solids analyses are in fair agreement with the values calculated from gas analyses shown in Figure J-35. Conversions range from 49 to 81% and show a trend of decreasing selectivity with increasing conversion.

Both the low conversions and the lower-than-equilibrium selectivities show that considerable opportunity exists for improving regenerator operation. One such improvement is a modified distributor which will give higher pressure drop and better fluidisation. A mound of agglomerated solids found around the distributor after each continuous run indicated that fluidisation had been incomplete. This mound decreased effective regenerator volume and probably caused poor contacting. Another possible cause of reduced regenerator efficiency was the presence of a vertical wall crack in the third run which could have allowed some gas by-passing. This crack will be repaired before the next run.

(3) Regenerator SO_2 Concentration

The time log of SO_2 concentration in regenerator off gas from the continuous runs appears in Appendix Figures J-5, J-13 and J-31. Concentrations of over 10% were obtained for brief periods, but most of the data from tests 1 and 3 were in the range of 6 to 8% SO_2. In test 2, the high limestone addition rate prevented achievement of high enough sulphur concentrations on lime to give good regeneration. From experience to date, a regenerator gas consistently containing 8% SO_2 appears to be an easily achieved goal. Even higher levels are possible but will require further improvements in regenerator design.

Figure 39 compares experimental SO_2 concentrations with the equilibrium curve. Points from alternate hours of pilot plant operations in Runs 1 and 3 are shown. Run 3 concentrations are somewhat lower than those from Run 1 and both sets of data fall well below equilibrium. Formation of sulphate in the cooler portion of the regenerator is responsible for the lower SO_2 concentrations. Apparently there is insufficient residence time and contacting in the hot portion of the regenerator to reach equilibrium by the solid-solid reaction, $CaS + 3CaSO_4 \longrightarrow 4CaO + 4SO_2$.

Changes which could improve selectivity of CaS oxidation and raise SO_2 concentration are:

(1) Increase regenerator volume.

(2) Preheat regenerator air.

(3) Make regenerator gas-solids flow counter-current.

Each of these changes would add to the cost of a CAFB regenerator, and whether or not they are justified remains to be established.

COMPARISON OF REGENERATOR SO_2 CONCENTRATION WITH EQUILIBRIUM CURVE

CAFB CONTINUOUS PILOT PLANT DATA

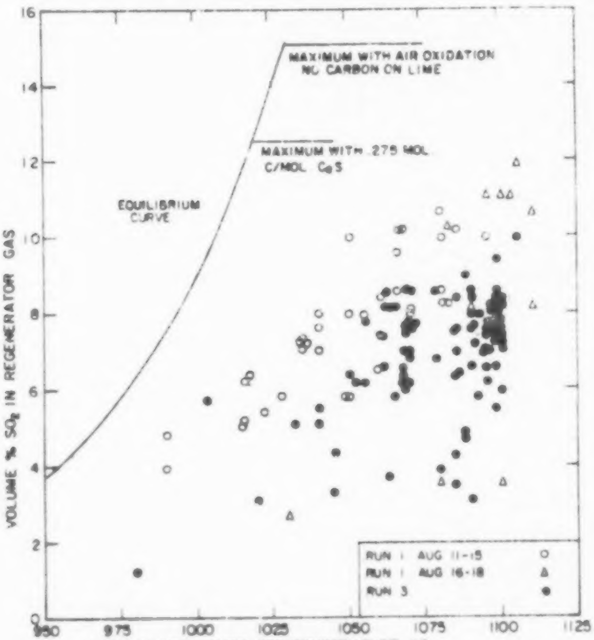

FIG 39

4. Vanadium Retention

Ability of the fluidised lime bed to absorb vanadium during fuel gasification is an important benefit offered by the CAFB process. If complete vanadium removal could be assured, boilers using CAFB could operate with higher superheater tube metal temperature and obtain improved efficiency without high temperature corrosion. The experience with vanadium retention has been variable. In the early studies, discussed in Appendix B, vanadium retentions ranging from 30 to 100% were observed. These early tests employed underfiring, that is, some fuel was burned beneath the distributor to increase fluidisation velocity at the bottom of the bed. Complete vanadium retention was obtained in these batch unit cyclic tests conducted during the limestone comparison portion of the current study. However, lower degrees of retention were measured in most cases during the cycle test variable study with stone BCR 1691. Appendix Tables E contain full data which are summarised here in Table XIV.

TABLE XIV

Summary of Vanadium Retention Results

Test	Average Gasification Temp. °C	Peak[a] Regeneration Temp. °C	Residence Time (sec)	Bed Loss Rate (g / min)	Make-up Rate (wt CaO/ wtS)	Potential Sulphur Differential (wt %)	Vanadium Loading (gV/ gbed/ cycle)	% Fuel Vanadium Removal
T1(C1toC14)	858	1024	0.15	1.4	2.4	3.0	0.000465	79
T3(C1toC14)	841	1019	0.15	2.0	2.5	3.0	0.000479	65
T3(C15toC23)	856	1034	0.22	1.6	2.8	2.6	0.00048	67
T3(C24toC31)	858	1046	0.15	2.8	2.3	3.6	0.000506	84
T3(C34toC41)	859	1028	0.15	0.1	3.1	3.1	0.000502	82
T4(C1toC10)	866	1038	0.15	3.4	2.2	3.2	0.000650	89
T5(C1toC20)	859	1017	0.21	2.9	5.1	2.2	0.000406	65
T6(C1toC17)	855	1022	0.21	4.5	2.5	2.2	0.000359	81
T8(C7toC20)	843	1008	0.24	6.0	2.5	2.0	0.000315	52
T8(C21toC34)	852	1020	0.24	2.5	1.7	2.2	0.000348	68
T8(C45toC61)	850	1015	0.28	5.2	2.7	1.7	0.000263	83
T8(C67toC80)	863	1046	0.15	2.1	2.2	2.2	0.000353	92
10	840	1026	0.36	7.1	2.7	1.5	0.000243	91
11	854	1017	0.25	3.8	2.1	2.2	0.000347	78

(a) Average of all cycles.

Regeneration temperature has given the most significant correlation with vanadium retention, but the correlation coefficient is only 0.6, so much of the variability remains unexplained. Figure 40 shows the variation of vanadium retention with peak regeneration temperature.

A hypothesis to explain the effect of regeneration temperature is that fixation of vanadium on the stone requires oxidising conditions and that kinetic limitations affect these fixing reactions. This hypothesis may also explain the apparent difference in temperature effect caused by underfiring.

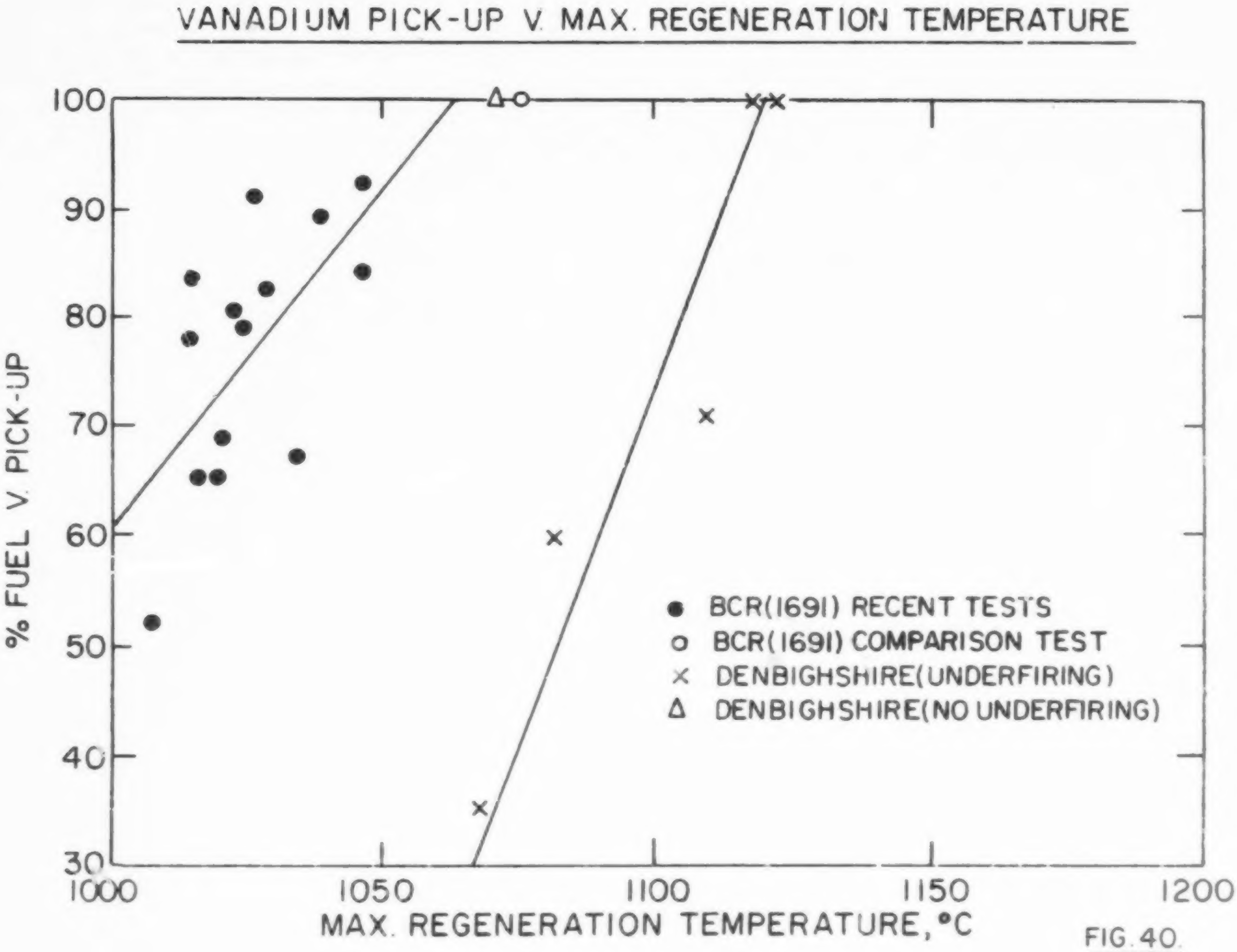
VANADIUM PICK-UP V. MAX. REGENERATION TEMPERATURE
% FUEL V. PICK-UP
100
90
80
70
60
50
40
30
1000
1050
1100
1150
1200
MAX. REGENERATION TEMPERATURE, °C
● BCR(1691) RECENT TESTS
○ BCR(1691) COMPARISON TEST
× DENBIGHSHIRE(UNDERFIRING)
Δ DENBIGHSHIRE(NO UNDERFIRING)

FIG. 40.

As Figure 40 shows, tests using underfiring required higher regeneration temperatures to obtain the same vanadium recovery as tests without underfiring. If vanadium is fixed under oxidising conditions there is greater opportunity for some of this fixation to take place during gasification if there is a strong oxidising zone near the distributor. With underfiring there is less oxygen in the air at the distributor and less driving force for oxidation.

The recent test series did not bear out the early observation that vanadium recovery correlates with iron pick-up. In these variable studies the iron content of stone has been constant. The earlier correlation, obtained under underfiring conditions, probably was due to the effect of temperature on both vanadium fixation and iron pick-up from metal parts of the unit.

Lime samples from the continuous pilot plant contain higher vanadium concentrations than can be accounted for by the vanadium content of the oil fed. Either there was selective loss of low vanadium fines or there was an extraneous source of vanadium. This matter will be investigated further in future runs.

5. Gasifier Product Gas Composition

Samples of gasifier product gas taken during the fresh bed test programme were analysed by gas chromatography. Although the GC method determines only the dry gas components, the missing quantities, H_2O and $>C_4$ hydrocarbons, were calculated from unit material balance considerations. Composition results were examined statistically and rationalised in the form of empirical equations which express the effects of variables on the yield of different compounds. The experimental results and the correlating equations are listed in Appendix D. Table XV, based on these equations, shows variations in gasifier product composition which can be expected over the normal range of temperatures and air/fuel ratios.

TABLE XV

Predicted Gasifier Gas Compositions

Air/Fuel Ratio:	15% of Stoichiometric		20% of Stoichiometric		25% of Stoichiometric		30% of Stoichiometric	
Bed Temperature:	850°C	900°C	850°C	900°C	850°C	900°C	850°C	900°C
Component								
CO_2 (% by vol.)	8.2	6.4	8.7	6.8	8.9	7.0	8.8	7.0
CO "	9.5	12.6	10.2	13.3	10.4	13.7	10.3	13.5
$C \leqslant C_4$ "	20.2	19.1	15.1	13.8	11.7	10.4	9.0	8.3
H_2 "	6.0	6.9	6.3	7.1	6.4	7.2	6.3	7.2
H_2O "	0.7	0.5	1.4	1.0	2.3	1.6	3.5	2.4
N_2 "	54.6	54.0	58.3	58.0	60.3	59.9	62.1	61.6
$C > C_4$ (%wt. of C fed)	20.1	21.7	19.6	21.9	15.6	17.5	10.2	10.6
CO/CO_2	1.16	1.97	1.17	1.95	1.17	1.96	1.18	1.93

Gas Velocity: 6 ft/sec. Bed Depth(Static): 24 inches.
Particle Size Range: 300-3175 µ.

This table shows how the carbon monoxide-dioxide ratio varies with temperature. At 850°C the ratio is close to unity whilst at 900°C it has increased to nearly two. A high ratio is desirable at low air/fuel ratios to increase the rate of carbon burn-off from the stone. As expected, the hydrocarbon content of the gas increases as air/fuel ratio is decreased, although it would appear that the greater-than-C_4 fraction reaches a maximum at around 20% of stoichiometric air. The water vapour content of the gas increases significantly as air/fuel ratio is increased, but the hydrogen remains fairly constant for a given temperature.

Increasing bed temperature from 850°C to 900°C, therefore, gives the following directional changes:

$$\begin{tikzcd}
CO_2 \arrow[r] & CO \\
 & CO \\
H_2O \arrow[ur] \arrow[dr] & \\
 & H_2 \\
 & H_2 \\
C \leqslant C_4 \arrow[ur] \arrow[dr] & \\
 & C > C_4
\end{tikzcd}$$

Increasing air/fuel ratio results in the preferential oxidation of the $> C_4$ fraction.

6. Lime Attrition and Losses

Loss of lime fines during continuous CAFB operation is undesirable and should be minimised. Although such losses will not be great enough under normal conditions to affect replacement lime requirements, they do constitute possible problems in boiler tube fouling and particulate emissions. The rate of lime loss in the continuous pilot plant depends both on gasifier cyclone efficiency and on rate of fines generation. However, the batch units do not return cyclone fines to the bed, so cyclone efficiency is not a factor. On the other hand, the batch units have expanded top sections which reduce fines losses by lowering gas velocity.

a. Cyclic Batch Test Bed Losses

Data on bed losses from the cyclic test variable study with BCR 1691 were analysed statistically and a correlation obtained with bed depth and lime replacement rate. The lined out loss rate, after seven cycles, was used in this correlation. Bed depth emerged as the most important variable with lime replacement second. Gas velocity had only minor effect over the range of 4 - 6 ft/sec studied. The regression equation obtained for the 6 ft/sec data was:

$$\text{loss rate, gm/min} = .00101\, m^{1.11} \quad h^{1.95} \tag{7}$$

where m is lime make-up rate, gm/min
h is bed depth, inches.

Figure 41 compares the test data with this correlation.

COMPARISON BETWEEN CALCULATED AND ACTUAL BED LOSS RATES(CYCLE TESTS)

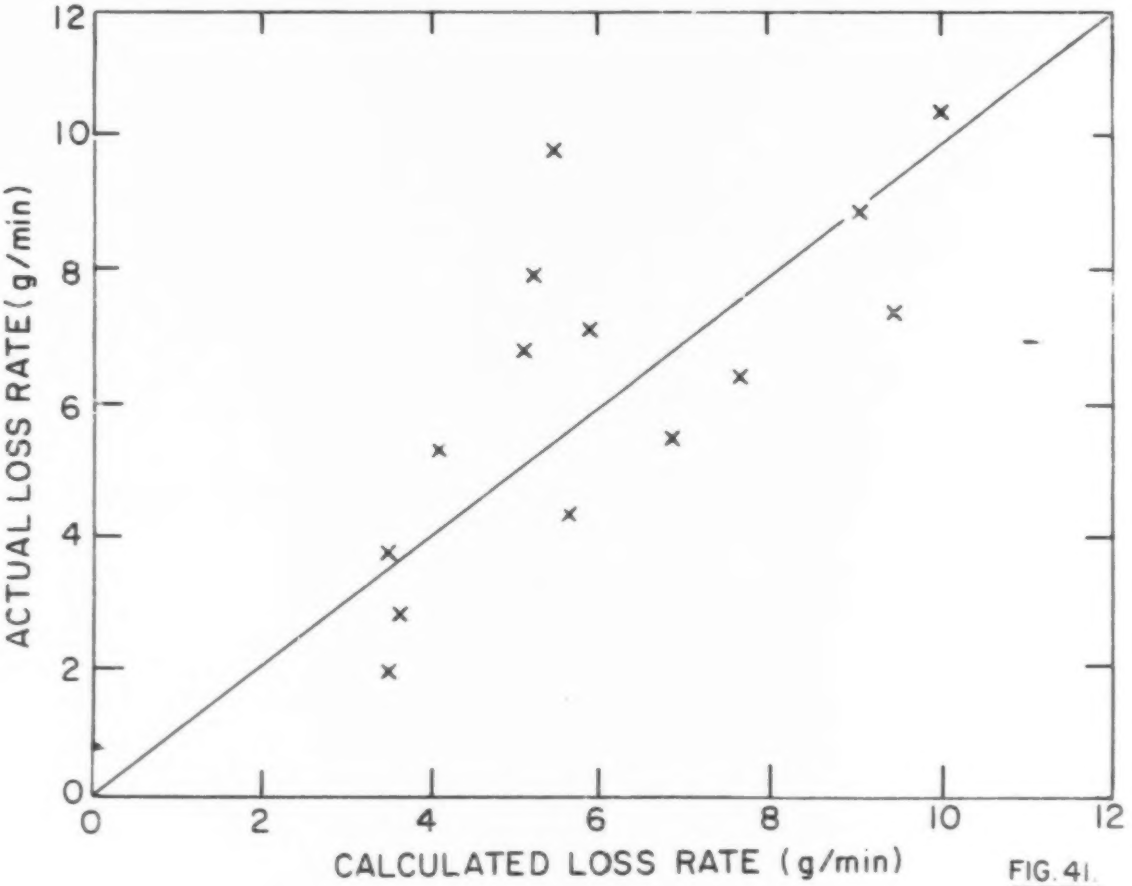

FIG. 41.

The magnitude of the bed depth effect is somewhat surprising. In part this increase in loss must stem from a higher attrition rate as bed volume increases with height. The decrease in disengaging height with increasing bed height also must play a part. But even with these factors, a 1.95 power effect of bed height on losses is unusually large for fluid beds. It is possible that operations were in a region where increase of bubble coalescence with bed depth caused a rapid change in the quantity of solids thrown violently into the gas space.

The nearly proportional effect of limestone make-up rate on bed losses is reasonable because of the decrepitation loss that occurs during in-situ calcination and because freshly calcined stone attrites **more** rapidly than aged stone. Evidence for the decrepitation loss was obtained during fresh bed calcinations where the total weight loss of the 300-3175 micron stone averaged 52.5%. The carbon dioxide loss due to calcination is only 37%. The higher loss rate of freshly calcined stone was shown by the fact that losses in the first cycle of a test series has been as much as seven times the lined out loss rate. It is not yet clear whether the change in loss rate with time is due to an increase in stone toughness or to an initial loss of a softer fraction.

b. Particle Size Changes

The variation of lime particle size distribution through a typical cycle test series with BCR 1691 at 6 ft/sec and 15.5 inch static bed depth is shown in Figure 42. In this test there was a trend toward increasing particle size with time. This trend is opposite in direction to that reported in the Interim Report(4) for a test of BCR 1691 at 4 ft/sec. Apparently the velocity difference was enough to increase the rate of fines loss faster than their rate of production. An opposite effect with increasing bed depth appears in Figure 43. Here we see that average particle size (measured after 20 cycles) decreased with increasing bed depth. The greater bed depth apparently increased the rate of fines production even more than the rate of fines loss.

c. Continuous Unit Experience

Lime losses in the batch unit at 6 ft/sec covered a range from 3 to 10 grams/min which is equal to 1.5 to 4.9 lb/hr-ft^2. Losses in the continuous pilot plant at 3 to 4 ft/sec. have tended to fall in the same range and have depended to a large degree on the efficiency of cyclone operations and in attrition produced by the oil feed inlets. The initial 40 hour period of the first pilot plant run had a loss rate of 1.44 lb/hr-ft^2. This increased to 2.7 lb/hr-ft^2 after the first decoking operation when one of the cyclone outlet tubes apparently was damaged. In the final hours of the test, rate increased to 9 lb/hr-ft^2 when neither cyclone apparently was operative.

VARIATION IN PARTICLE SIZE DISTRIBUTION, TEST 3

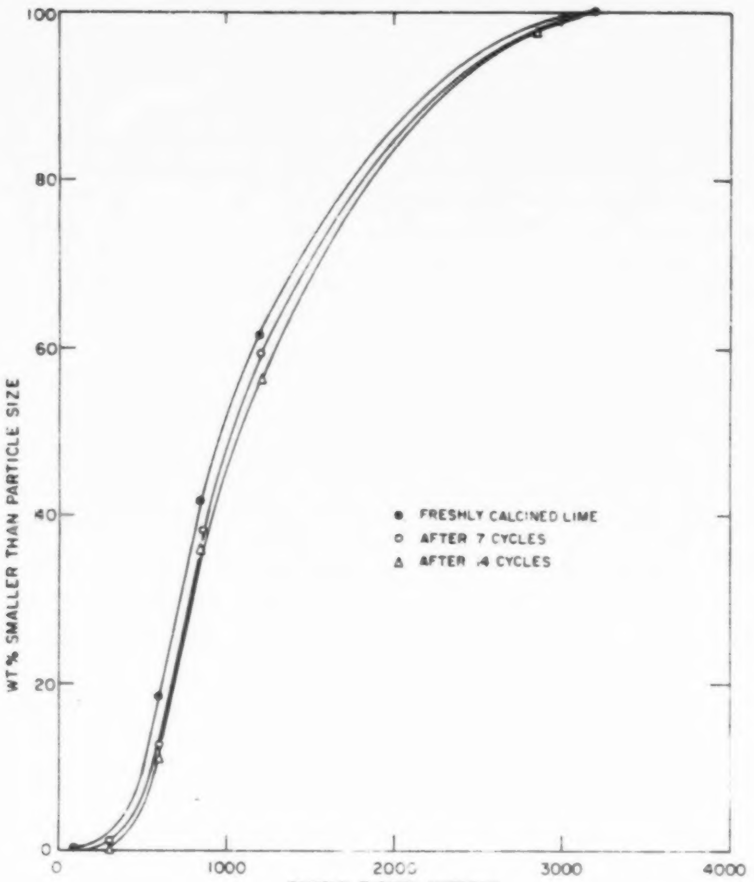

FIG. 42

CHANGES IN MEAN PARTICLE SIZE WITH BED DEPTH (AFTER 20 CYCLES)

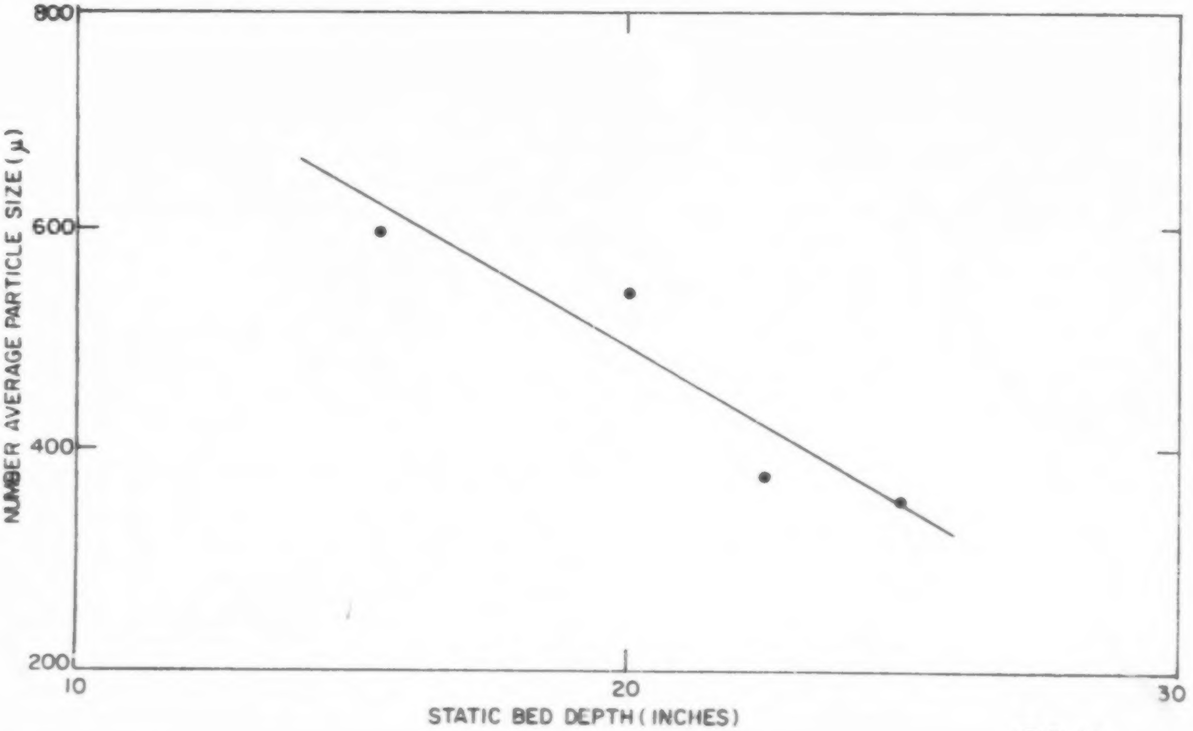

FIG. 43.

The second pilot plant test is not considered to be representative of any practical operating condition. Two factors in that test combined to produce entirely unacceptable lime loss rates. These were elimination of cyclone gas outlet dip tubes altogether and use of high gas velocities [illegible] the fuel distributors. This high gas velocity (above 400 ft/sec) was intended to improve fuel atomisation and distribution (which it did but, in addition, it gave severe lime attrition. Losses from the bed in this test ranged from 6 to 14 lb/hr-ft^2.

The final test run used cyclone gas outlet tubes which could be steam-cooled during decoking, and, in addition, employed a controlled temperature decoking procedure to avoid damaging the tubes. Nevertheless, in this test too, cyclone operation failed during a portion of the test when solids that had accumulated on material test specimens in the gas transfer duct were dislodged and blocked the cyclone solids drain lines. For the first 65 hours of operation, lime losses from the bed averaged 1.25 lb/hr-ft^2. During the test period between 130 and 181 hours while cyclones were inoperative, the loss rate reached a high level of 3.7 lb/hr-ft^2.

This level of lime loss, even with good cyclone operation, is higher than desired. It certainly would require particulate removal equipment before the flue stack and could affect the rate of solids accumulation on boiler surfaces. We suspect that unnecessary attrition is taking place in the vicinity of the air distributor nozzles and have designed a new distributor which will reduce the inlet jet velocity while maintaining sufficient pressure drop to provide stable operation.

The method of fresh limestone addition in the pilot plant contributed to the high lime losses also. The stone was merely dropped through a connection in the lid. Though simple, this procedure caused the fresh stone to fall through the rising gas stream which undoubtedly carried a portion of it straight through the outlet duct. The pilot plant is being modified to use a low level pneumatic lime feeder which will avoid this source of dust losses.

7. Operation of Continuous Pilot Plant

Three runs were made in the continuous pilot plant in 1971. It was the intention in each of these runs to operate, if possible, for a period of 200 hours of gasification, but the first two tests were interrupted short of this goal. The third run lasted 230 hours of which 204 were gasification. The unit was shut down voluntarily at the end of the third run.

Appendix J contains details of the continuous run log of operations, results, and inspections following runs. Figures start on page 291.

a. Gasification and Sulphur Removal

The principal objective of the CAFB process:fuel gasification with sulphur removal,was demonstrated throughout the three tests. The product gas from the gasifier ignites readily and burns with a bright luminous stable flame. Smoke free operation with about 1.5% oxygen in the boiler gas has been demonstrated for long periods. This is better performance than found with normal fuel oil firing in this boiler which requires about 3% oxygen in the flue gas.

The variation of sulphur removal efficiency with time is shown in Appendix Figure J-16. Ability to achieve 100% sulphur removal was demonstrated in a number of instances. In most cases, this high removal efficiency was reached with lime replacement rates of over 1.5 times stoichiometric, but removal of over 95% was reached with less than stoichiometric lime addition during the final day of the third run.

b. Solids Circulation

Blockages in circulation of solids between beds were experienced in the first run, but this problem was eliminated in runs 2 and 3. Figure J-30 shows Run 3 circulation rates as estimated from heat balance. Higher circulation rates have been required than originally estimated because the extent of calcium sulphate formation in the reagent has been greater than expected. The sulphate forming reaction releases extra heat and requires more solids circulation for regeneration temperature control.

Circulation problems in the first run were due to obstructions that formed in the transfer line where air was injected into beds of defluidised sulphided solids. It had been intended to use flue gas in the bed transfer pulse injector system. However, the system for cleaning and compressing the flue gas proved unreliable due to plugging by wet lime fines and air had been substituted. At points where air entered the defluidised bed of sulphided lime there was partial regeneration which appears to have resulted in momentary formation of a liquid sticky phase and agglomeration of the solids.

These blockages did not occur during periods of steady operation in Run 1 but formed during interruptions to operations which occurred for other reasons.

The problem was eliminated in Runs 2 and 3 by use of nitrogen as the injection gas for the bed transfer system. No circulation blockages were experienced in 91 hours of continuous operation in Run 2 or in 230 hours of operation in Run 3. In Run 3 there were several shutdowns and restarts for decoking, and in each case bed recirculation was resumed without difficulty. At the end of Run 3 a small deposit of agglomerated lime fines was found in the regenerator to gasifier transfer duct. This deposit had not obstructed flow, but whether it was stable in size or would have continued to grow is not certain. Similarly, there were accumulations of agglomerated fines at the entries to the solids transfer down comers both in the gasifier and the regenerator. Neither of these accumulations had obstructed flow.

In cold circulation tests with the continuous pilot plant mockup it had been observed that the bed transfer system was able to move approximately two pounds of lime per cubic foot of air. The transfer system in the hot pilot plant was considerably more efficient with rates as high as 25 pounds per cubic foot of nitrogen being achieved.

c. Gasifier Temperature Control

The gasifier temperature was adjusted by varying the proportion of flue gas recycle to fresh air in the fluidisation gas. This adjustment was made by the operator in response to the reading of the bed temperature controller. While other conditions were stable, this manual control was adequate to hold temperature within about 5 Deg. C of target. However, when other factors, such as limestone or oil feed rate, varied, there could be greater fluctuation of temperature. Even so, it was possible to hold the temperature between 870 and 900°C which appears to be adequate for the process. Automation of the reactor temperature control would be a simple matter, but the cost of the controller and valves does not appear to be justified at present.

d. Regenerator Temperature Control

An automatic controller for the regenerator temperature had been installed prior to the first run but was not used until the third run. In the first two runs, the temperature was varied by manual adjustment of the solids transfer rate. In the third run this adjustment was turned over to the automatic controller with excellent results. Control could be maintained within plus or minus 5 Deg. C of the set point, and response to set point changes was rapid. The controller was able to compensate for minor changes in pressure differential between gasifier and regenerator. Thus solids circulation rate was more constant with the temperature controller in operation than it had been under manual control. The improvement due to the controller can be seen by comparing uniformity of regenerator temperature (Figures J-4, J-12, and J-29) before and after 2300 on the first day of Run 3 when the automatic system was activated.

e. Limestone Feed

A simple hopper with a vibrated transfer line leading to an opening in the gasifier lid was used to feed limestone to the gasifier. The vibrator on the transfer line was turned on and off by a pressure switch operating on gasifier level. Solids feeding was interrupted when it became necessary to refill the hopper. A stream of air was admitted with the stone to prevent backflow of gas to the feeder.

This system was not quite satisfactory and a modified system is under study. Disadvantages of the old system were:

(1) Interruption of stone feed to fill the hopper affected gasifier temperature;

(2) Dropping the stone through the gasifier lid increased dust losses;

(3) Feed rate from the vibrator was variable and sensitive to gasifier pressure;

(4) Instantaneous feed rate could not be determined, only averages over the periods required to empty a hopper.

In general the feed system required more operator attention than is desirable. Between Runs 2 and 3 a pneumatic feed line had been installed to inject limestone directly into the bed. However, the vibrator used to feed solids from the hopper to the transfer line did not prove to be satisfactory, and this system was not used. An improved pneumatic feeder will be used in the next test.

f. Refractory Condition (1)

Details of the unit condition at the end of Run 3 appear in Appendix J. In general the unit has stood up well. Some initial roughening of the refractory in the bed region occurred during cold circulation trials prior to the first run but little additional errosion has been observed. Several vertical cracks have appeared at the ends of the gasifier. There is also a vertical crack in the lower portion of the regenerator. It appears that these cracks reseal themselves in the upper, hot portion of the unit during gasification.

(1) End of run photographs start on page 326.

It is probable that a crack remains open in the cold plenum region of the gasifier and of the regenerator. The gasifier air distributor is fixed to a closed metal plenum so that no bypassing is possible there. Some bypassing through the crack may have occurred in the regenerator. A new metal distributor fixed to the plenum has been designed for the regenerator to eliminate possible bypassing in future tests.

No penetrations between gasifier and regenerator spaces have appeared.

The interior cyclone surfaces have been roughened to a considerable degree. At the end of each run the cyclone surfaces have also shown a rough accumulation of lime and carbon scale. It appears that maintaining a smooth interior cyclone surface will be difficult.

The surfaces of solids transfer ducts have remained smooth and intact.

g. Agglomerations and Deposits

Agglomerates and solid deposits have been encountered at several points in the unit during the three runs. The cause of some of these deposits has been identified and the deposits eliminated. Others remain as potential operating problems.

(1) Regenerating Solids

The blockage of solids transfer lines in Run 1 has been mentioned. This blockage was part of a more general problem area. It appears that regeneration of sulphided lime produced a transient liquid state in the CaO - CaS - $CaSO_4$ system and that the liquid formed can cement lime particles together if they are in close proximity in a static bed. This transient liquid phenomena has been observed by Consolidation Coal Co.(5) in their work on a CO_2 acceptor process. In Run 1 there were several instances where air was introduced into a hot static bed of defluidised sulphided lime. In each of these instances a lump of agglomerated lime particles resulted. In subsequent runs it has been possible to avoid conditions leading to this agglomeration in most cases. The only persistent problem area has been the zone directly above the regenerator distributor.

In each of the three runs there has been accumulation of an agglomerated lump directly above the distributor. This agglomerate has been attributed to local poor fluidisation due to insufficient regenerator distributor pressure drop and to the top-hat distributor design which creates a defluidised centre region. A new regenerator distributor has been designed to alleviate this problem.

A slight deposit of cemented fines has been observed in the regenerator gas space. At the end of Run 1 the deposit was located just below the gas outlet. At the end of Run 3 there was no deposit in the regenerator itself, but a small amount was found in the outlet duct at the expansion joint. It is possible that there is some critical temperature at which fines will adhere to the regenerator wall and that the location of this temperature zone depended on operating conditions. In Run 3 the regenerator temperature was higher and more uniform than in Run 1. The quantity of this upper regenerator deposit was small and is not considered a serious problem.

(2) Gasifier Air Distributor

In the first two pilot plant runs there had been negligible increase of pressure drop through the gasifier air distributor. The third run however encountered a steady pressure rise that lasted throughout the operating period. The initial pressure drop of 9.5 inches w.g. had increased to 33.8 in. by the end of the run. Inspection of the distributor after the run revealed that all of the nozzles had some degree of obstruction, 20% of the holes were at least 25% obstructed, and about 2% were completely blocked. The obstructions appeared to consist of fine particles that had adhered to the periphery of the nozzle bores. There fines probably came from the recycle flue gas.

The occurrence of distributor blockage in the third run but not the first two is somewhat surprising as the quantity of fines in the recycle flue gas was much less in the third run than in the others. A cyclone had been installed in the flue gas returning line after the second run. It had been effective in removing a great deal of solids as the flue gas lines and the plenum itself were relatively clean after the third run but had contained much solids after the first two. Fines content of the recycle gas had been particularly high in Run 2. A possible explanation for the greater adhesion of fines to the distributor nozzles in Run 3 is the higher level of sulphur on gasifier lime. In its passage through the boiler this sulphur was partly converted to sulphate. Whether or not this sulphate would have rendered the lime particles sticky at conditions in the distributor nozzles remains to be established.

New distributor nozzles are being fabricated for the gasifier to reduce lime attrition due to the distributor jet velocity. Since these nozzles will retain small orifices and will still be subject to plugging by fines, a cyclone of increased efficiency will be installed in the flue gas recycle line.

(3) Carbon in Gasifier Outlet

Deposition of carbon and lime on surfaces of the gas outlet system appears to accompany gasification. In all of the tests to date there has been a gradual build up of pressure drop between gasifier and boiler during gasification due to this deposition of solids. Figures J-3, J-11, and J-25 show the increase in total gasifier pressure for the three runs. Of course the total pressure drop from gasifier to atmosphere is the sum of several pressure drops. The breakdown of these various pressure drops appears in Figure J-27 for Run 3.

The initial pressure drops from gasifier to cyclone outlet and from cyclone outlet to the boiler are approximately equal.

Both pressure drops increased during each gasification period until the total pressure drop forced a decoking operation to remove the cause of obstruction. In Run 1 the decokings were somewhat uncontrolled and the stainless steel cyclone gas outlet tubes were destroyed. In Run 3 a controlled decoking procedure was used and the cyclone outlet tubes remained intact. The procedure used in decoking was to first stop fuel addition and go to a very high flue gas recycle rate for fluidisation until the sulphide of the lime bed was converted to sulphate. Then the bed was slumped while a flue gas-air mixture was added above the bed to burn out carbon deposits in the cyclone inlets, cyclone bodies and tubes, and bifurcated duct between cyclones and burner. When carbon burning was complete, gasification was resumed. In some cases it was necessary to go to full combusting conditions to raise or hold bed temperature before resuming gasification. Steam cooling was used to control the temperature of the cyclone outlet tubes during decoking in addition to the reduced oxygen atmosphere provided by flue gas recycle.

The solids laid down in the gas outlet system are a mixture of lime and carbon. Burning removes the carbon but leaves a loosely agglomerated layer of lime. Therefore decoking alone is not completely adequate and some form of soot blowing probably will be required in a commercial installation. In the third pilot plant run the solids were dislodged by rodding out through the duct at 1600 hours of day 6. These solids appear to be soft and readily removed with slight mechanical force.

There are insufficient data yet to establish a correlation, but it appears reasonable that conditions of low air/fuel ratio and low gasification temperature which increase the concentration of heavy hydrocarbons in the gasifier product will increase the rate of deposition in the outlet ducts. On the other hand, operation at temperatures over 900°C, as on Day 5 of the third run, did not reduce the rate of pressure increase in the cyclones. Neither did local injection of air into the cyclone inlet ducts.

At the end of Run 1, the gasifier was shut down without a burn out so that solids accumulations could be observed in place. Large obstructions were found in the cyclone inlets which, at that time, were square entry rectangular ducts. Between the first and second runs, the entry sections of these ducts were rounded to a more stream-lined profile to reduce entry loss pressure drop. This measure appears to have helped as pressure rise in the third run was much slower than in the intact cyclone tube period of Run 1.

The point of transition from the vertical cyclone outlet tube to the horizontal gas duct represents another sharp change of gas direction where solids accumulation is high. The outlet system is being modified to provide a rounded transition at this point.

The problem caused by solids accumulation in the gas ducts was aggravated in the early runs by the relatively low head available from the gasifier blower system. Although two blower stages had been installed, the intention had been to use the first stage only as a booster during startups. Therefore the flue gas recycle line had been piped to the inlet of the second stage. With this arrangement the flow rate of fluidising gas fell as back pressure on the gasifier increased. This reduction had the effect of decreasing air/fuel ratio and causing carbon laydown and back pressure to increase even faster. The piping system was revised during a decoking period on Day 6 of the third run to return flue gas to the first stage blower and permit use of both blowers in series. The result was much greater operating flexibility and the ability to extend run length between decokings.

The effect of carbon deposition on gas passage pressure drops is more severe in the case of a pilot plant than it will be in a commercial unit with large size ducts. On the other hand, the rate of accumulation of solids per unit of duct surface probably is comparable. Thus the rate of pressure rise in a large unit will be much lower than the pilot plant has experienced. The decoking frequency of a large unit will probably be dictated by limitations on the quantity of carbon to be burned in each decoking rather than by pressure drop limitations.

(4) Boiler Tube Fouling

A large portion of the lime that passed through the gasifier cyclones was deposited in the boiler at the end of the main firetube and at the inlets to the first pass of $1\frac{3}{4}$ inch fire tubes. In Run 2 a large quantity of solids was removed from drains at the boiler back end and the small tubes were only slightly obstructed at their inlets. However, in Run 3 there was a progressive increase of pressure drop across the boiler throughout the run, though less solids were withdrawn through the drains. The rate of pressure drop increase accelerated following the rodding of the duct on day 6 and it is assumed that the cyclone dust return lines became blocked during that rod out. Blockage of these lines rendered the cyclones inoperative and increased the rate of lime loss from the gasifier to the boiler. The rate of solids removal from the boiler back end and flue gas final cyclone increased sharply following this rodding as the solids recovery curves in Figures J-19 and J-21, show.

The rate of pressure drop increase accelerated also as the diameters of the tube inlets became progressively smaller.

In several cases the fire tube inlets were completely obstructed by a crusty coating of fines. The crust could be broken easily leaving a core shaped deposit in many of the tubes. These cores penetrated for a short distance into the tubes but did not extend the entire length. The longest core was 15 inches long with the majority being less than 6 inches. The core material was approximately the composition of calcium sulphate.

These fire tube deposits originate from highly sulphided lime fines which convert partially to sulphate and become sticky during their passage through the flame.

The deposits observed are not hard and could be removed easily when the boiler had cooled. It may be possible to remove these deposits under hot operating conditions by soot blowing procedures, and tests of this possibility are planned for the future. A deposit measuring probe will be inserted in the boiler for the next test. Nevertheless, deposits are undesirable, and their rate of formation should be minimised. Obvious steps to be taken are better maintenance of cyclone operability to prevent excessive loss of lime fines and reduced attrition to reduce rate of production of fines. A third means of reducing deposits may be to reduce stickiness of the lime by decreasing the sulphur level on gasifier solids. This step may be accomplished by improving regenerator efficiency to increase both selectivity and conversion level in regeneration. Such an improvement also should reduce the lime replacement required for good sulphur removal efficiency.

(5) Test Specimens

A closed tube of silicon nitride was suspended in the bifurcated duct above the left hand cyclone outlet in Run 2. A second silicon nitride specimen was placed above the right hand cyclone and a self bonded silicon carbide specimen above the left hand cyclone in Run 3. The silicon nitride specimen was removed intact at the end of Run 2, but it developed cracks soon after and displayed extreme brittleness. The silicon nitride specimen in Run 3 was completely destroyed during the rodding out operation on Day 6. It too had evidently become very brittle as no portions of it were recovered when cyclone fines were examined after cyclone draining.

The self bonded silicon carbide specimen was removed intact after 104 hours of operation. Examination by the manufacturer revealed that its strength and toughness had not deteriorated during the test.

It had been subjected to three decoking operations. On the basis of this test, arrangements have been made to install cyclone gas outlet tubes of self bonded silicon carbide before the next pilot plant trial. Durability of the material therefore can be evaluated under realistic operating conditions.

D. Conclusions and Recommendations

This Phase I study of the CAFB process has resulted in the following conclusions and recommendations.

1. Conclusions

a. The Venezuelan fuel oil originally selected by NAPCA (now OAP) performs in a satisfactory manner in the CAFB process.

b. U.S. Limestone BCR 1691 is suitable for CAFB applications. It has only 81% (initial sample) to 89% (large supply) of the CaO content of the U.K. stone formerly used as a standard, but at a high enough replacement rate it can give nearly complete sulphur removal from the fuel.

c. The tested sample of U.S. Limestone BCR 1690 was inferior to the U.K. stone standard and to BCR 1691. It contained only 60% as much CaO as the original sample of BCR 1691, it suffered greater attrition, it gave lower sulphur removal efficiency even when replaced at equal CaO/S rates, and it tended to sinter and agglomerate under CAFB cyclic conditions.

d. Gasification temperature and air/fuel ratio interact in their effect on sulphur recovery in a CAFB gasifier. As gasification temperature decreases, high air/fuel ratios must be employed to assure adequate removal of carbon from the lime and avoid rapid deactivation toward sulphur absorption. Gasification temperatures above 870°C permit air/fuel ratios as low as 16% of stoichiometric with good sulphur removal. Gasification temperature as low as 800°C can be used with air fuel ratios above about 27% of stoichiometric.

e. Sulphur removal efficiency of a CAFB gasifier begins to decline at temperatures above 900°C, possibly because of partial regeneration in the vicinity of the air distributor.

f. The sulphur removal efficiency of fresh lime beds improves as average particle size decreases. At low levels of lime utilisation the lime utilisation at equal efficiency is roughly inversely proportional to the square of number average particle diameter. In cyclic tests on a continuous unit, consideration also must be given to the effect of particle size range on losses from the bed, especially during in-situ calcination of the stone. A limestone with lower particle size of 300 microns was less effective overall than one with 600 microns minimum size because of the high loss rate of the former.

g. The effects of gasifier bed depth and fluidisation on sulphur removal efficiency are correlated by their ratio, the superficial gas residence time above the fuel injector. As an approximation, sulphur removal follows a first order rate expression.

h. Lined out sulphur removal efficiency with the large supply of U.S. limestone BCR 1691 was lower in batch cyclic tests than predicted by the performance of the original sample in the limestone screening tests. Greater limestone replacement rate was required to achieve the same sulphur removal efficiency. In part this effect was due to greater losses of the low particle size end of the 300-3175 micron stone selected for the variable study compared with 600-1400 micron stone used in the comparison test.

i. Potential Sulphur Differential, the quantity of sulphur (expressed as wt. % on lime) to which the lime charge was exposed in each gasification cycle, emerges as an important variable in the batch unit cyclic variable study. The sulphur removing reaction rate appears to vary inversely with the 1.9 power of this differential. The empirical equation which correlates the results of the batch cyclic tests is:

$$\%\ \text{Sulphur Removal Efficiency} = 100 \left[1 - \frac{1}{\text{EXP}\left(\frac{9.3\ t^{.6}\ m^{.95}}{d^{1.9}}\right)} \right]$$

where t = gas residence time, seconds, in the bed above the fuel injector.

m = lime replacement rate, wt. CaO/wt.S

d = % sulphur differential.

This relationship implies that beds in the region of 3 to 4 ft deep will be required for >90% sulphur removal with stoichiometric lime replacement or less and gas velocities of 6 ft/sec.

j. Heat to raise lime temperature from the gasification level to the regeneration level is supplied in part by oxidation of CaS and in part by oxidation of carbon on the lime. At temperatures below 1000°C most of the CaS oxidation produces $CaSO_4$. Therefore the lower the gasification temperature the greater is the amount of $CaSO_4$ formed and retained by the lime after regeneration. Low gasification air/fuel ratio increases the carbon content of the lime and reduces the amount of sulphate formation. The quantity of sulphate formed in batch unit regenerations also increases with gas velocity and with average lime particle size.

k. The pilot plant regenerator required temperatures 50 to 100°C higher than predicted by equilibrium considerations to achieve a given level of oxidation selectivity to calcium oxide. Contacting and kinetic limitations appear to be responsible for the increased temperature requirement. Temperatures of 1070 to 1100°C gave SO_2 concentrations of up to 10% in the regenerator off gas.

l. Vanadium retention in the batch unit cyclic test series correlated with temperature in the regenerator-higher regeneration temperatures giving more complete vanadium retention. Essentially 100% retention of vanadium was achieved with regeneration temperatures above about 1065°C.

m. Vanadium concentrations in solids removed from the continuous pilot plant gasifier and regenerator beds were several times higher than expected for uniform vanadium deposition. These data imply selective loss of low vanadium fines and accumulation of vanadium concentrates.

n. Composition of gasifier product varies with temperature and air/fuel ratio in a manner that can be expressed by empiriral equations. Increasing the temperature increases CO/CO_2 ratio, raises the hydrogen and greater-than-C_4 fraction, and decreases water formation. Raising the air-fuel ratio gives preferential oxidation of the greater-than-C_4 fraction.

o. Lime losses in batch unit studies increased rapidly with bed depth but were little affected by velocity in the 4 to 6 ft/second region. Loss rates of 1 to 9 lb/hr-ft^2 were experienced. The continuous pilot plant encountered lime fines losses ranging from 1 to 14 lb/hr-ft^2 with cyclone operation being the major controlling factor and jet attrition by air admitted with the fuel being a contributing factor.

p. The continuous pilot plant functions as intended by producing a hot, combustible low sulphur fuel gas from heavy fuel oil while delivering a concentrated stream of SO_2 from the regenerator. A 230 hour pilot plant run was completed in which 204 hours were at gasifying conditions. This test employed 2.5% S Venezuelan fuel oil and Denbighshire limestone. Sulphur removal from the fuel exceeded 99% for long periods, and 95% removal was achieved with less than stoichiometric lime replacement.

q. Introduction of oxygen containing gas into unfluidised hot beds of sulphided lime causes agglomeration of the particles, presumably by formation of a transient liquid state in the CaS - $CaSO_4$ - CaO system undergoing regeneration.

r. The solids circulation system incorporated in the continuous pilot plant performs in a satisfactory manner. It delivered 230 hours of trouble-free operation during the third pilot plant run prior to planned run termination.

s. Automatic regenerator temperature control by regulation of solids circulation rate performs in a satisfactory fashion.

t. Gasifier temperature control by flue gas recycle is feasible and produces satisfactory temperature regulation.

u. Lime particles in flue gas recycle can cause gradual blockage of the gasifier air distributor.

v. Gradual accumulation of a deposit of carbon and lime in cyclones and ducts leading from the gasifier to the burner forces periodic decoking operations. Periods up to 64 hours between decokings have been demonstrated in the pilot plant to date with cyclones functioning and longer periods without cyclone outlet tubes. Decoking by controlled burnout has been demonstrated by a procedure which requires about two hours of interruption to gasification.

w. A portion of lime fines from the gasifier adheres to the entry sections of the pilot plant boiler tubes causing a gradual increase in pressure drop through the boiler. The scale is not hard and was easily removed following the 230 hour pilot plant run.

x. Operability and flexibility of the CAFB pilot plant indicate that it is a useful tool for study of the CAFB process under realistic operating conditions and that it will be capable of providing answers to a majority of CAFB process and engineering questions.

2. Recommendations

a. Continued use of the Venezuelan fuel oil of the type studied to date is recommended for the next stage of CAFB pilot plant studies.

b. Use of limestone BCR 1691 in the next stage of pilot plant studies is recommended.

c. Gasification temperatures in the range of 870 to 900°C are recommended as offering the most favourable range for the majority of CAFB studies.

d. A limestone feed with a particle size range of 600 to 3200 microns is recommended for the majority of experimental work in the next phase of pilot plant studies.

e. It is recommended that greater bed depths and gas velocities be explored in the next phase of pilot plant work to minimise bed area and lime replacement requirements for a given CAFB duty.

f. An improved air distributor is recommended for the regenerator to avoid local areas of defluidisation and to improve regenerator contacting performance.

g. Frequent sample analysis of all solids streams for vanadium are recommended in future pilot plant studies to establish a vanadium balance.

h. A redesign of the gasifier air distributor is recommended to reduce local jet velocity and thereby to decrease lime attrition rate.

i. An independent source of inert, oxygen free gas, is recommended for use at any CAFB point where gas is injected into hot sulphided solids which may be defluidised. The pilot plant now uses nitrogen, but a commercial installation could employ an inert gas generator. Dependence on flue gas recycle is not recommended as it may be subject to interruption when most needed.

j. A more efficient solids separator is recommended for the flue gas recycle line to reduce possibility of plugging the gasifier air distributor by recycled lime fines.

k. Close observation of the quantity of carbon laid down in gas transfer ducts is recommended in future pilot plant tests to establish conditions which moderate carbon laydown and to permit prediction of carbon deposition under commercial conditions.

l. Provision of facilities for controlled decoking of transfer ducts is recommended in commercial CAFB units.

m. Trial of self bonded silicon carbide as a material of construction for pilot plant cyclone gas outlet tubes is recommended.

n. Installation of a probe in the end of the pilot plant boiler is recommended for study of the rate and nature of solids deposits to be expected on boiler superheater tubes.

o. Major process features of the CAFB process have been demonstrated, and possible operating problem areas having been defined, it is recommended that the continuous CAFB pilot plant be used in an intensive study of CAFB process variables and engineering features to solve those problems and provide data needed for design of a 150 megawatt boiler conversion.

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APPENDICES
TO
FINAL REPORT

OAP CONTRACT CPA 70-46

STUDY OF CHEMICALLY ACTIVE FLUID BED GASIFIER FOR REDUCTION OF SULPHUR OXIDE EMISSIONS

APPENDIX	SUBJECT
A	CAFB Chemistry
B	Results of work Just Prior to Contract
C	Batch Unit Fuel and Lime Comparisons
D	Batch Reactor Variable Study with Fresh Lime Beds
E	Batch Reactor Cycle Test Variable Study
F	Correlation Functions for Sulphur Removal Efficiency
G	Design Basis for the Pilot Plant Gasifier System
H	CAFB Pilot Plant Alarm Systems
I	CAFB Pilot Plant Operating Procedures
J	Operation of Continuous Pilot Plant
K	Burning Rate Model for Carbon on Lime in CAFB Gasifier

APPENDIX A

CAFB CHEMISTRY

Chemistry of CAFB Gasification Process

In gasification, sulphur compounds react with lime under reducing conditions. In contact with hot lime particles (800-900°C) the fuel oil vaporises and cracks to liberate compounds such as H_2S, COS and CS_2. Sulphur is then absorbed by the lime by reactions such as those listed in Table A-I.

TABLE A-I

EQUILIBRIUM CONSTANTS AND HEATS OF REACTION FOR DESULPHURISING REACTIONS DURING GASIFICATION

	Reaction	$\log_{10}K_p$ 800°C	$\log_{10}K_p$ 900°C	$\log_{10}K_p$ 1000°C	$\log_{10}K_p$ 1100°C	ΔH (kcal/mol) at 900°C
1	$CaO + H_2S \rightleftharpoons CaS + H_2O$	2.97	2.91	2.82	2.70	-4.7
2	$CaO + COS \rightleftharpoons CaS + CO_2$	4.70	4.42	4.13	3.83	-17.9
3	$CaO + \frac{1}{2}CS_2 \rightleftharpoons CaS + \frac{1}{2}CO_2$	4.91	4.63	4.28	3.90	-19.7

The equilibria for these reactions are well to the right indicating sulphur partial pressures of less than 200 ppm up to 1100°C (2000°F) as shown in Figure A-1

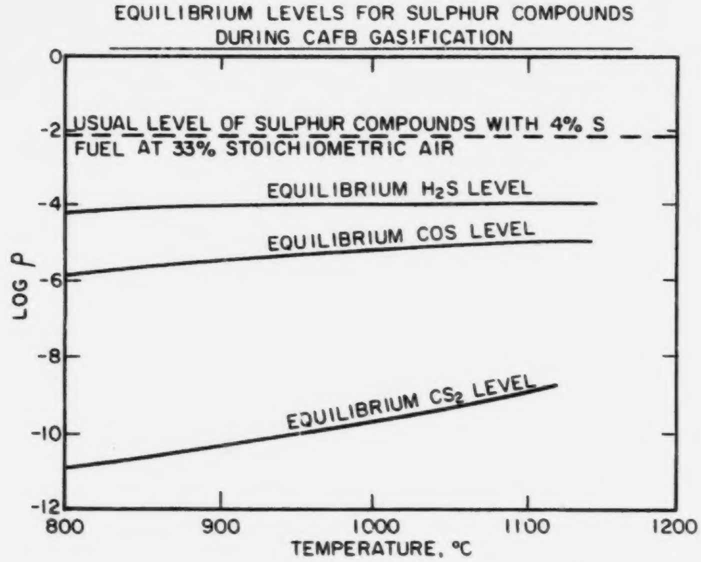


Figure A-1

With a fuel containing 4% sulphur these equilibria imply a desulphurising efficiency of greater than 90% up to 1100°C. However other factors limit the effective gasification temperature to the 800-900°C range.

In the shallow fluidised bed of the gasifier there is rapid circulation of lime between top and bottom. There is evidence that coke is laid down on the lime in the upper portion of the fluid bed by oil cracking and coking reactions and that this coke burns off in the lower portion, where oxygen is supplied by the air distributor.

Only a portion of the CaO in the lime is reacted on each pass of solids through the gasifier. Good sulphur absorption reactivity has been obtained with up to 20% of calcium reacted in laboratory tests, but in the continuous unit the average extent of reaction will only be to about 10% of the calcium in the bed.

The lime reagent is regenerated by oxidation of its calcium sulphide content. Again, several reactions may occur as shown in Table A-II.

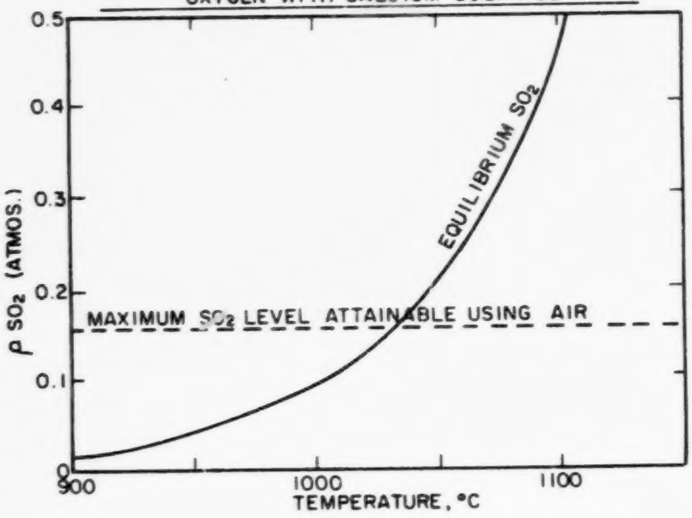
TABLE A-II

EQUILIBRIUM CONSTANTS AND HEATS OF REACTION FOR THE REGENERATION OF SULPHIDED LIME

Reactions		$\log_{10} kp$					ΔH (kcal/mol)
		900 °C	1000 °C	1100°C	1200°C	1300°C	at 1100° C
4	$CaS + 1\frac{1}{2}O_2 \rightleftharpoons CaO + SO_2$	16.11	14.43	13.01	11.87	10.86	-109.5
5	$CaS + 2\,O_2 \rightleftharpoons CaSO_4$	23.96	20.61	17.79	15.46	13.46	-220.2
6	$CaS + 3\,CaSO_4 \rightleftharpoons 4CaO + 4SO_2$	-7.44	-4.09	-1.36	1.09	3.12	221.5

Both reactions (4) and (5) are highly exothermic and their equilibria under conditions of excess oxygen are far to the right. The SO_2 partial pressures to satisfy their respective equilibria at different temperatures are shown in Figure A-2. (These are also the equilibrium SO_2 partial pressures for reaction (6)).

Maximum SO_2 partial pressure with air regeneration should, therefore, be obtained at temperatures in excess of 1040°C (1900°F) when the reaction is conducted at atmospheric pressure.



EFFECT OF TEMPERATURE ON THE EQUILIBRIUM SO_2 PARTIAL PRESSURES FOR THE REACTION OF OXYGEN WITH CALCIUM SULPHIDE

Figure A-2

APPENDIX B

RESULTS OF WORK JUST PRIOR TO CONTRACT

This appendix summarises results obtained by the Esso Research Centre on the CAFB gasifier in the period between submission of the Technical Proposal (3) to NAPCA and signing of the contract. These studies were carried out with the UK fuel and Denbighshire limestone whose properties are listed in Table III of the text. Batch fluid bed reactors of 7-in.i.d. were used in these tests.

A. Reagent Life

The effect of successive adsorption/regeneration cycles on reagent activity during partial combustion was studied at different bed make-up rates and mean particle sizes. Absorption was carried out at a stoichiometric air rate between 30 and 35%, a superficial gas velocity of 2.5 ft/sec. and a bed temperature between 800 and 900° C. In each adsorption period the bed was exposed for a total of 100 min [illegible] during which its sulphur content rose to between 3 and 4%. Regeneration was carried out by shutting off the fuel supply and allowing air alone to contact the bed. Bed samples were taken for sulphur analysis at the beginning and end of each adsorption period, and a regular check was made on bed losses; fresh material being added to the bed after each cycle to make up for attrition and sampling losses. Bed make-up rate was controlled by adjusting the number and size of the samples. Varying mean particle size in a downward direction only was considered since an increase would be unlikely to give a beneficial effect. This was achieved in the low gas velocity batch fluidiser by fitting baffles above the bed which gave better retention of the smaller particles and also by changing the size distribution of the limestone feed to 600-1200 μ from the original 850 - 1200 μ. These alterations gave a reduction in mean particle size of lime in the bed from 810 to 635 μ.

Figure B-1 illustrates the effect that successive adsorption/ regeneration cycles can have on reagent activity.

These results show that, under the particular operating conditions employed, the desulphurising efficiency dropped from 100% to a fairly steady level of 60% and that this steady level was reached in less than 10 cycles. The rapid drop in efficiency is typical of all tests where a steady level of less than 100% was observed.

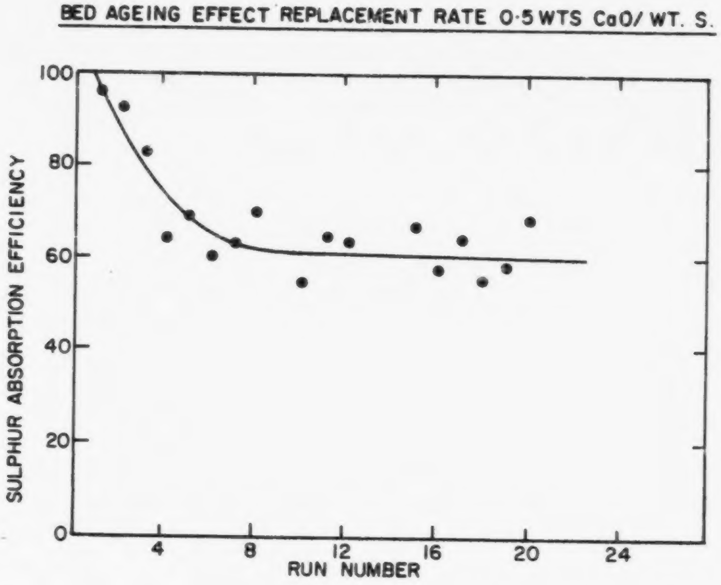


FIGURE B-1

Figure B-2 summarises the results obtained in tests on the effect of bed make-up rate and mean particle size.

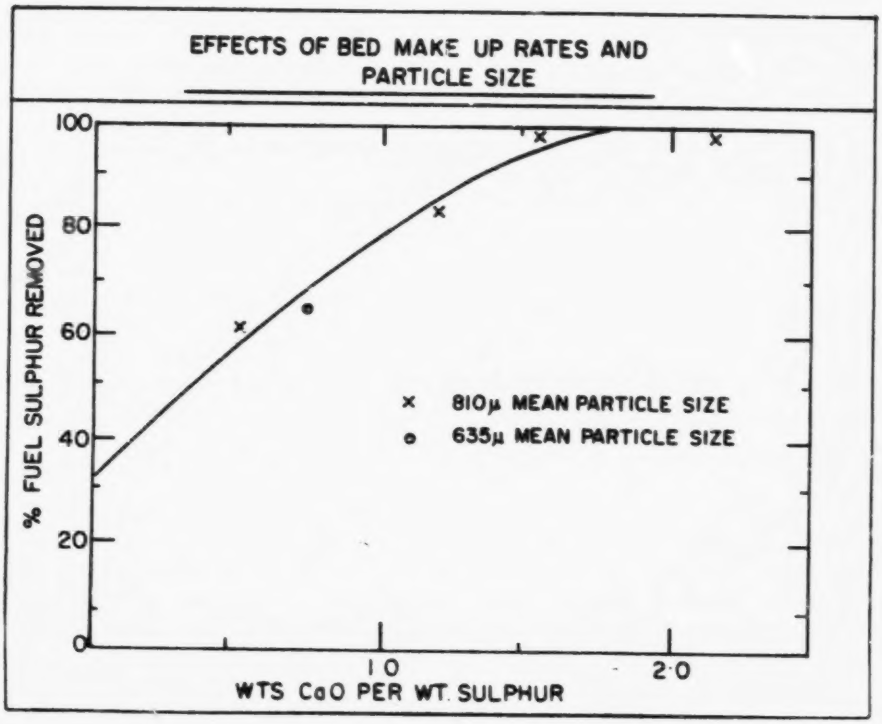


FIGURE B-2

These show that at bed make-up rates in excess of 1.8 wts. CaO per wt. of sulphur, a desulphurising efficiency of 100% can be maintained indefinitely and that at lower make-up rates, the efficiency falls off rapidly. Reducing the mean particle size had no significant effect on the steady level of desulphurising efficiency within the limited range covered.

B. Retention of Sodium and Vanadium by the Bed

During the life tests, the retention by the bed of sodium and vanadium present in the fuel oil was studied.

Sodium pick-up was found to be consistently low, being less than 20% of the amount introduced.

The results obtained on vanadium retention were more encouraging but also more diverse. Under conditions of partial combustion, the efficiency of vanadium pick-up varied from test to test between the levels of 35% and 100% and was not related to bed make-up rate (Figure B-3).

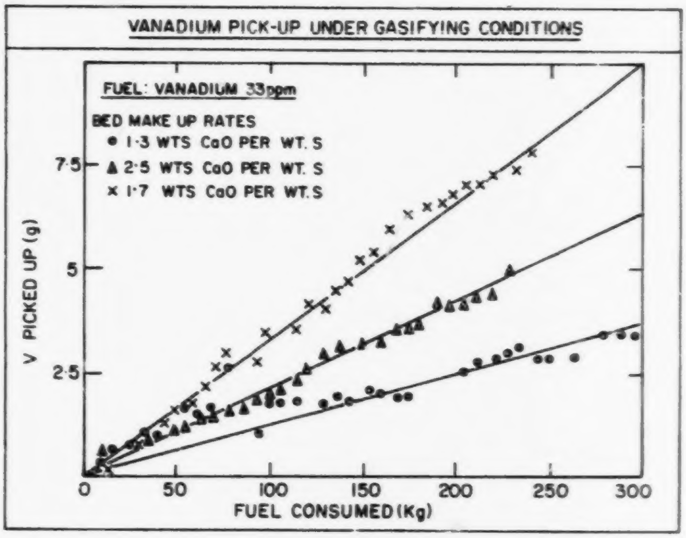


FIGURE B-3

A correlation was found to exist however between rate of vanadium pick-up and rate of iron pick-up (Figure B-4). The iron originated from the metal components of the unit which protrude into the bed.

Subsequent work during the contract has indicated that the iron-vanadium correlation is not one of cause and effect but that vanadium retention and iron pickup were independently related to regeneration temperature.

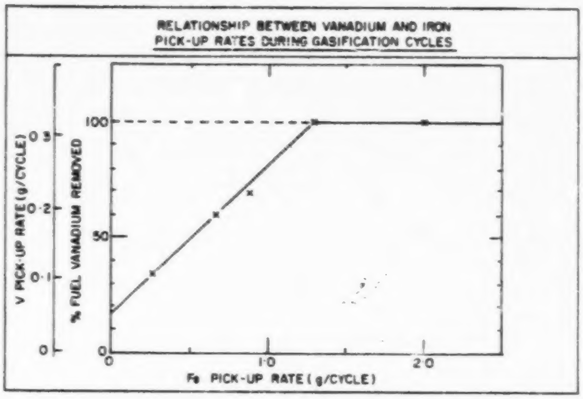


FIGURE B-4

C. Gasification without Underfiring

All gasification tests previously reported had been carried out in the low gas velocity batch fluidiser at superficial gas velocities of less than 3 ft/sec. At these velocities, the combustion of kerosene below the distributor was necessary to preheat the fluidising air and so prevent the deposition of coke at and below the level of the fuel injectors. The utilisation of 25% of the incoming air in this way had resulted in the sulphur being evenly distributed throughout the bed as calcium sulphide. Tests carried out early in 1970 in the tall, No.3 batch reactor showed that underfiring was unnecessary when superficial gas velocities in excess of 3 ft/sec were employed. Operating without underfiring almost certainly prolongs distributor life and, for this reason alone, it is the preferred mode of operation. However, it also makes the bed more complex chemically in that there is a transition from calcium sulphate to calcium sulphide with increasing distance from the distributor and, along with this, an increase in total sulphur content. These effects result from the additional oxygen entering through the distributor. This oxygen reacts with the calcium sulphide below the level of the fuel injectors and at the temperature of the absorbing bed (800-900°C) predominantly forms calcium sulphate. The sulphur dioxide which is released during the less favourable formation of calcium oxide is reabsorbed as calcium sulphide above the level of the fuel injectors.

The effect of this change in operation on the variables affecting desulphurising efficiency has been studied to some extent. Results so far show that an operating temperature between 800 and 900°C is still to be preferred. However, the influence of stoichiometric air rate has changed radically as is evident in Figure B-5 where the effects of stoichiometric air rate on the desulphurising efficiency of fresh beds are compared with and without underfiring.

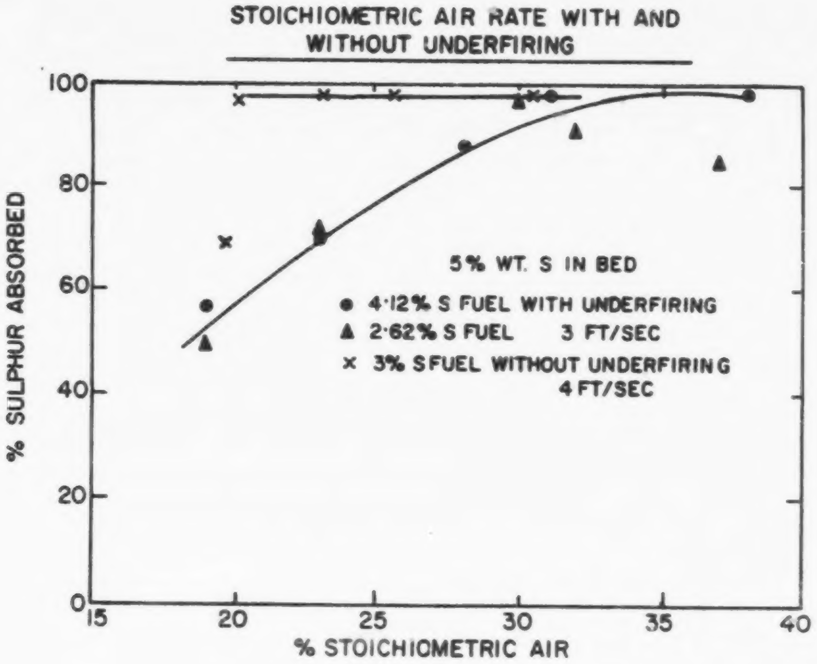


FIGURE B-5

Figure B-6 shows the manner in which desulphurising efficiency varied with increasing CaO utilisation during the course of these high velocity runs. Except for the run at 20.7% of stoichiometric air, the lime maintained its activity to over 20% of utilisation.

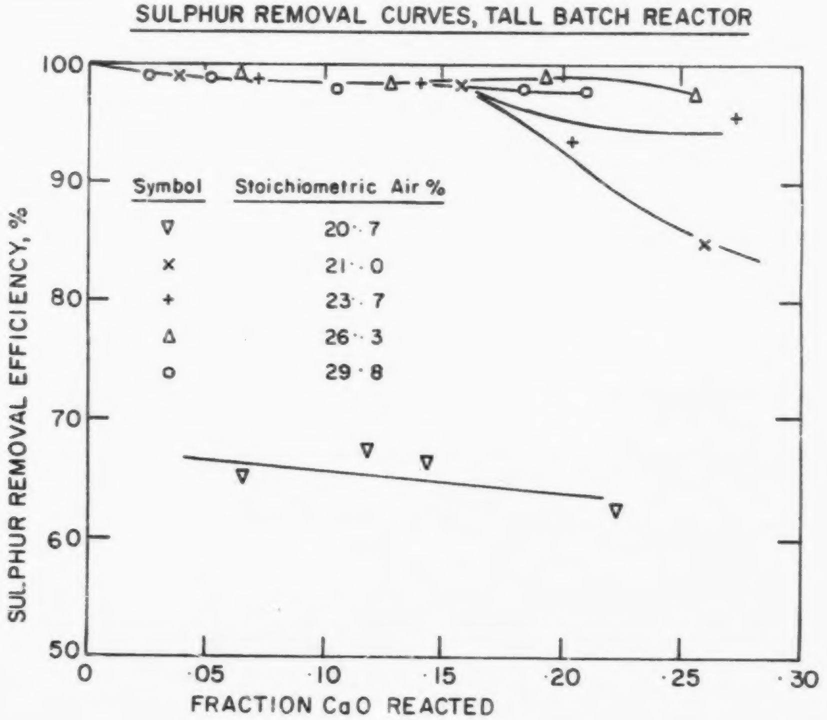
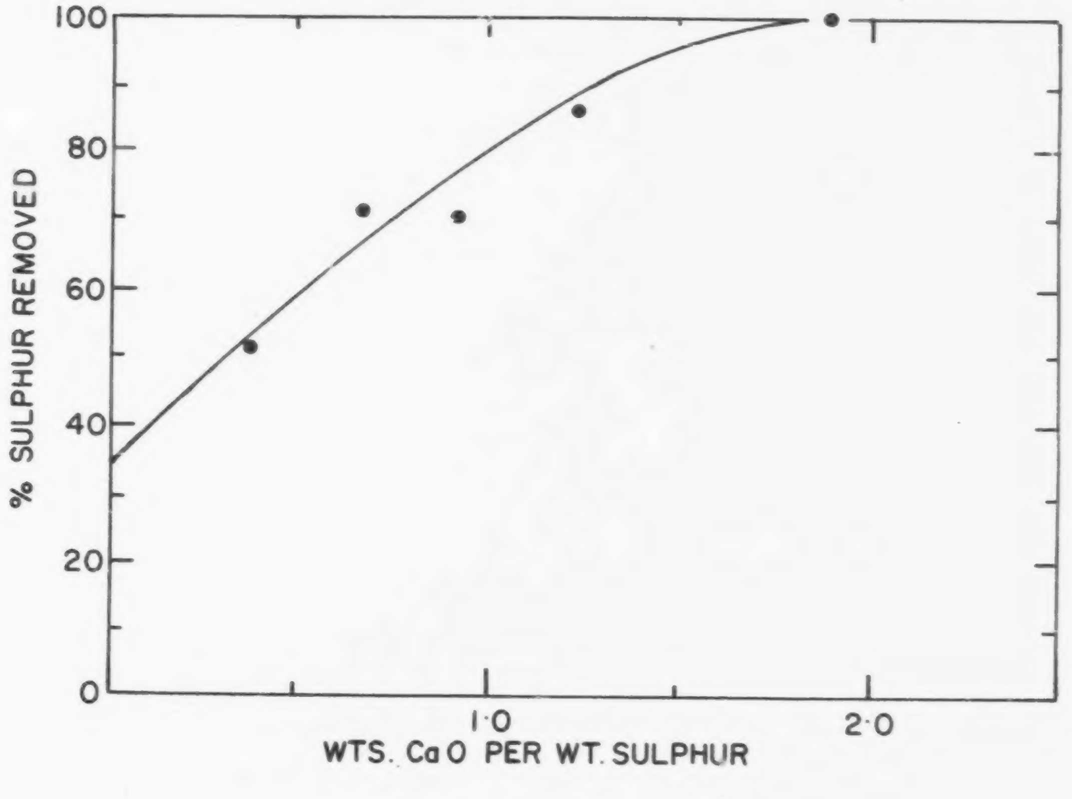


FIGURE B-6

It is evident that with no underfiring high desulphurising efficiency can be maintained at lower stoichiometric air rates. If it is assumed that the drop in efficiency previously observed with reductions in stoichiometric air rate was due to the observed build-up of carbon on the particles, then it is likely that the current improvement is due to an increased amount of carbon being removed from the bed in the oxidising zone resulting in a continual supply of carbon-free stone to the gasifying region.

To be able to operate at the lower stoichiometric air rate and higher gas velocity is a great advantage since it means that the fuel throughput is increased for a fluid bed of given cross-sectional area.

Multicycle tests reveal no significant effect of the absence of underfiring on the relationship between lime replacement rate and the lined-out efficiency of sulphur removal. The line established in the absence of underfiring, Figure B-7, is quite similar to that found with underfiring, Figure B-2. The data in Figure B-7 were obtained at 25% of stoichiometric air compared with 30-35% used with the underfiring tests in Figure B-2.



EFFECT OF MAKE-UP RATE ON DESULPHURISING EFFICIENCY

FIGURE B-7

APPENDIX C

BATCH UNIT FUEL AND LIME COMPARISONS

Appendix C contains experimental data from batch reactor studies in the 1-A unit during preliminary limestone and fuel comparison studies.

TABLE C-1

SHEET 1 of 6

CAFB BATCH UNIT TEST CONDITIONS

1-A Unit Preliminary Fuel and Limestone Comparisons

Run No.	Fuel	Limestone	Limestone Particle Size µ	Fuel Rate (g/min)	Air Rate (L/min)	Superficial air vel. (ft/sec.)	% Stoich. Air	Temperature C Initial	Temperature C Average
11(c)	UK(a)	UK	800 - 1200	200	473	4.3	21.7	770	865
22(c)	"	"	"	150	441	4.0	27.0	790	840
23(c)	US(b)	"	"	150	435	3.9	26.6	790	830
25(c)	UK(a)	"	"	200	440	3.9	20.2	780	835
26	"	"	"	195	450	4.1	21.0	830	840
27	"	"	"	200	443	4.0	20.3	850	840
28A	"	"(d)	"	200	464	4.2	21.3	840	840
28B	"	"(d)	"	200(e)	459	4.2	21.1	860	850
28C	"	"(d)	"	200(f)	420	3.8	19.3	790	840
29	"	"(d)	"	200	480	4.4	22.0	915	860
30	"	"(d)	"	200	471	4.3	21.6	730	870
32	"	UK(d)	"	200	434	4.0	19.9	900	840

(a) U.K. fuel 3.03% S
(b) U.S. fuel 2.22% S
(c) Fuel injected to bed centre
(d) Used bed of lime employed in test
(e) 1.5 CFM air to fuel nozzle
(f) 0.5 CFM air to fuel nozzle (other tests use 1.0 CFM air to fuel nozzle)

TABLE C-1

SHEET 2 of 6

CAFB BATCH UNIT TEST CONDITIONS

1-A Unit Preliminary Fuel and Limestone Comparisons

(Cycle Test, UK Fuel and Limestone)

Run No.	Cycle	Air Rate 1/min.	Superficial Gas Vel. ft/sec	Fuel Rate g/min	% Stoich Air	Feed Particle Size µ	Bed Depth (Static) in.	Limestone Make-up Rate g/cycle	Average Bed Temp. (Absorption) °C	Max. Bed Temp. (Regeneration) °C
33	1	426	3.9	165	23.7	850-1400	15.5	400	860	1050
34	2	432	3.9	"	24.0	"	-	"	880	1060
35	3	439	4.0	"	24.4	"	-	"	870	1080
36	4	448	4.1	"	24.9	"	15.4	"	870	1070
37	5	444	4.1	"	24.7	"	-	"	880	1080
38	6	442	4.0	"	24.6	"	15.6	"	890	1070
39	7	442	4.0	"	24.6	"	-	"	890	1080
40	8	423	3.9	"	23.5	"	15.2	"	880	1070
41	9	424	3.9	"	23.6	"	-	"	900	-
42	10	412	3.8	"	22.9	"	-	"	880	-
43	11	423	3.9	"	23.5	"	15.7	"	880	1070
44	12	415	3.8	"	23.1	"	-	"	870	1060
45	13	438	4.0	"	24.3	"	15.2	"	880	1070
46	14	414	3.8	"	23.0	"	-	"	880	1090
47	15	414	3.8	"	23.0	"	-	"	890	1080
48	16	413	3.8	"	23.0	"	-	"	890	1070
49	17	424	3.9	"	23.6	"	15.2	"	880	1060
50	18	422	3.9	"	23.5	"	-	"	880	1080
51	19	418	3.8	"	23.2	"	-	"	880	1070
52	20	426	3.9	"	23.7	"	-	"	880	1070
53	21	419	3.8	"	23.3	"	15.2	"	880	1070

60 minute absorption cycle

Normal, Low pressure drop, distributor

TABLE C-1

SHEET 3 of 6

CAFB BATCH UNIT TEST CONDITIONS

1-A Unit Preliminary Fuel and Limestone Comparisons

(Cycle Test UK Fuel and US Limestone 1691)

Run No.	Cycle	Air Rate l/min.	Superficial Gas Vel. ft/sec	Fuel Rate g/min	% Stoich Air	Feed Particle Size µ	Bed Depth (Static) in.	Limestone make-up Rate g/cycle	Average Bed Temp. (Absorption) °C	Max. Bed Temp. (Regeneration) °C
54	1	441	3.9	165	24.5	600-1400	16.5	-	840	1050
55	2	447	4.0	165	24.8	"		500	850	1060
56	3	461	4.2	165	25.6	"		500	865	1060
57	4	411	3.8	165	22.9	"	16.0	500	880	1060
58	5	422	3.9	165	23.4	"		500	875	1070
59	6	414	3.8	165	23.0	"		500	875	1070
60	7	408	3.8	165	22.7	"		500	855	-
61	8	419	3.8	165	23.3	"	16.1	500	855	-
62	9	433	4.1	165	24.1	"		500	875	1090
63	10	415	3.8	165	23.1	"		500	885	1080
64	11	415	3.9	165	23.1	"		500	885	1080
65	12	419	3.9	165	23.3	"		500	885	1080
66	13	416	3.8	165	23.1	"		500	870	1080
67*	14	394	3.6	165	21.9	"	16.1	500	860	1070
68	15	413	3.8	165	23.0	"		500	865	1080
69	16	411	3.8	165	22.8	"		500	870	1090
70	17	407	3.7	165	22.6	"		500	865	1080
71	18	403	3.7	165	22.4	"		500	870	1070
72	19	394	3.7	165	21.9	"		500	880	1080

60 minute absorption cycle

Normal, Low pressure drop, distributor

* 2nd attempt at cycle. In 1st attempt fuel pump failure after 21 min and spontaneous regeneration.

SHEET 4 of 6

TABLE C-1

CAFB BATCH UNIT TEST CONDITIONS

1-A Unit Preliminary Fuel and Limestone Comparisons
(Cycle Test UK Fuel and US Limestone 1690)

Run. No.	Cycle	Air Rate l/min.	Superficial Gas Vel. ft/sec	Fuel Rate g/min	% Stoich Air	Feed Particle Size μ	Bed Depth (Static) in. Start Cycle	Bed Depth (Static) in End Cycle	Limestone make-up Rate g/cycle	Average Temp. (Absorption) °C	Max. Bed Temp. (Regeneration) °C
78	1	399	3.6	165	22.2	600-1400	15.5	13.4	-	845	1080
79	2	412	3.7	"	22.9	"	15.6	-	1800	850	1060
80	3	413	3.7	"	23.0	"	-	-	"	830	1060
81	4	408	3.7	"	22.7	"	-	-	"	845	1070
82	5	431	3.7	"	24.0	"	-	18.1	"	840	1070
83	6	412	3.7	"	22.9	"	16.9	-	"	855	1090
84	7	434	4.0	"	24.1	"	-	-	"	860	1080
85	8	473	4.3	"	26.3	"	-	-	"	870	1080
86	9	430	3.9	"	23.9	"	-	-	"	865	1080
87	10	441	4.1	"	24.5	"	-	16.8	"	870	1050
88	11	396	3.6	"	22.0	"	16.1	-	840	860	1060
89	12	432	4.0	"	24.0	"	-	-	"	880	1070
90	13	422	4.0	"	23.5	"	-	-	"	890	1070
91	14	422	4.0	"	23.5	"	-	-	"	890	1070
92	15	420	3.9	"	23.3	"	-	15.5	"	890	1080
93	16	385	3.5	"	21.6	"	15.9	-	"	845	1070
94	17	385	3.5	"	21.6	"	-	-	"	830	1060
95	18	400	3.7	"	22.2	"	-	14.2	"	865	1060

60 minute absorption cycle

Normal, Low pressure drop, distributor

TABLE C-1

CAFB BATCH UNIT TEST CONDITIONS

SHEET 5 of 6

1-A Unit Preliminary Fuel and Limestone Comparisons

Run No.	Fuel	Lime-Stone	Limestone particle size µ	Fuel Rate g/min	Total Air Rate l/min	Superficial Gas Vel. ft/sec.	% Stoich. Air	Bed Depth (in) (Static) Start	End	Average Bed Temp. (Absorption °C	Max. Bed Temp. (Regeneration) °C	Duration of Absorption (min.
96	UK(3.05%S)	UK	600-1400	210	441	3.9	19.3	15.5	14.2	800	1090	300
98	"	"	"	208	433	3.9	19.1	"	14.4	815	1080	126
99	"	"	"	195	435	3.9	20.5	"	14.2	810	1070	120
100	"	"	"	245	540	4.8	20.2	"	13.8	810	1090	90
101	"	"	"	184	444	4.1	22.1	"	14.2	850	1090	120
102	"	"	"	175	435	4.1	22.8	"	14.6	860	1090	120
103	US(2.22%S)	US(1691)	"	165	446	4.2	24.8	"	14.2	865	1090	240
104	UK(3.05%S)	UK	"	173	440	4.1	23.3	"	14.4	860	1080	120
105	"	"	"	165	433	4.0	24.1	"	14.2	845	1070	120
106	"	"	"	142	454	4.2	29.3	"	14.2	860	1050	205
107	"	"	"	206	449	3.8	20.0	"	-	740	1000	60
108A	"	"	"	206	453	3.9	20.2	-	-	770	-	30
108B	"	"	"	165	456	4.1	25.4	-	14.4	820	1010	30
109(a)	"	"	"	206	445	4.0	19.8	15.5	15.6	810	-	120

High pressure drop distributor

(a) Run ended without regeneration, cooled under nitrogen blanket

TABLE C-1

CAFB BATCH UNIT TEST CONDITIONS

1-A Unit Preliminary Fuel and Limestone Comparisons

SHEET 6 of 6

Run No.	Fuel	Lime-Stone	Limestone Particle Size Range µ	Fuel Rate (g/min	Total Air Rate (1/min	Superficial Gas Vel. ft/sec.	% Stoich. Air	Bed Depth (in) (Static Start	End	Average Bed Temp.(Absorption) °C	Max. Bed Temp.(Regeneration) °C	Duration of Absorption (min.)	Distributor
110	UK.3.03%S	UK	600-1400	207	401	3.5	17.7	15.5	14.2	810	1090	129	Normal
111	"	"	"	190	430	3.8	20.7	15.5	14.0	820	1085	120	"
112	"	"	"	161	449	4.1	25.2	15.5	14.4	855	1090	124	"
113	"	"	"	163	433	3.8	24.2	15.5	14.0	825	1060	125	"
114	"	"	"	176	415	3.6	21.5	15.5	14.1	815	1065	122	"
115	"	"	"	150	419	3.8	25.6	15.5	14.4	850	1090	123	"
116	"	"	"	204	429	3.6	19.3	15.5	[illegible].3	780	1055	122	High Pressure Drop
117	"	"	"	205	444	3.8	19.9	15.5	14.0	790	1060	120	"
118	"	"	"	190	400	3.5	19.3	15.5	14.2	815	1080	120	"
119	"	"	"	205	423	3.6	18.9	15.5	14.2	790	1070	120	"
120	"	"	"	173	420	3.7	22.3	15.5	14.4	830	1080	120	"
121	"	"	"	150	436	3.9	26.7	15.5	14.3	850	1060	120	"

SHEET 1 of 3

TABLE C-2

SULPHUR REMOVAL IN FRESH BED BATCH REACTOR TESTS

% OF FEED SULPHUR REMOVED

1-A Unit

Run No. Time, Mins.	11	22	23	25	26	27	28-A	28-B	28-C	29	30	32
15	77.3	100	100	86.8	-	-	-	-	-	-	-	-
30	79.3	-	-	62.3	-	75.5	-	-	-	-	-	83.7
35	-	95.0	-	-	-	-	-	-	-	-	-	-
45	-	95.2	-	64.1	-	-	-	-	-	-	-	-
60	-	95.1	-	64.8	-	65.4	-	-	-	-	-	74.4
75	-	-	93.7	61.6	-	-	-	-	-	-	-	-
90	83.2	94.7	92.7	64.5	95.7	-	-	-	-	-	-	-
105	-	-	-	61.5	-	-	-	-	-	-	-	-
120	-	90.4	-	60.9	86.8	-	-	-	-	-	-	-
135	-	-	-	-	-	-	-	-	-	-	-	-
150	82.7	84.7	92.0	-	75	60.2	-	-	-	-	-	-
180	-	86.1	-	-	-	-	-	-	-	-	-	-
210	83.8	-	84.1	-	-	58.4	-	-	-	-	-	-
240	-	-	-	-	-	43.4	-	-	-	-	-	-
270	80.7	-	79.1	-	-	-	-	-	-	-	-	-
300	-	-	76.3	-	-	-	-	-	-	-	-	-
330	-	-	78.3	-	-	-	-	-	-	-	-	-
E10(a) %	82.5	92.5	90.5	61	97.5	62.5	-	-	-	-	-	-
Carbon Rate g/min	-	.15	.69	7.83	2.22	2.84	.82	.89	1.27	.20	.31	6.68

(a) E10 = Sulphur removal efficiency at 10% CaO reacted

TABLE C-2

SHEET 2 of 3

SULPHUR REMOVAL IN FRESH BED BATCH REACTOR TESTS

% FUEL SULPHUR REMOVED

1-A Unit

Run No: Time (min)	96	98	99	100	101	102	103	104	105	106	109
10	88.5	-	99.1	99.4	100	100	100	100	100	100	100
20	87.6	72.7	96.8	97.6	84.2	99.2	100	89.4	100	100	-
30	84.5	74.6	91.1	90.4	81.0	99.3	100	87.1	88.3	100	-
40	82.6	74.6	86.2	85.9	79.7	99.5	100	86.6	86.6	100	-
50	82.4	74.3	86.0	83.1	83.2	99.5	100	86.3	86.4	100	84.6
60	82.7	72.7	85.9	83.3	82.6	92.8	100	86.6	-	99.1	-
70	77.8	72.7	85.1	83.2	82.6	92.3	100	86.8	-	98.3	-
80	73.9	74.2	84.5	83.3	81.9	92.1	100	86.4	84.2	97.5	-
90	75.4	73.9	83.6	84.3	81.3	90.5	94.7	86.0	83.8	96.8	80.7
100	76.9	73.5	83.2	-	81.0	90.2	91.0	85.7	87.3	96.8	-
110	78.4	72.7	83.6	-	82.2	89.1	89.1	85.3	85.3	96.7	78.8
120	75.7	71.6	83.6	-	81.1	88.8	88.0	85.0	85.3	96.2	-
130	74.6	72.7	-	-	-	-	86.8	-	88.7	95.8	-
140	72.6	-	-	-	-	-	86.2	-	-	-	-
150	73.4	-	-	-	-	-	84.2	-	-	-	-
160	73.5	-	-	-	-	-	82.7	-	-	-	-
170	69.4	-	-	-	-	-	81.7	-	-	-	-
180	-	-	-	-	-	-	82.2	-	-	-	-
190	76.6	-	-	-	-	-	80.9	-	-	92.9	-
200	69.8	-	-	-	-	-	81.1	-	-	91.9	-
210	78.7	-	-	-	-	-	80.4	-	-	91.3	-
220	76.1	-	-	-	-	-	79.9	-	-	-	-
230	72.6	-	-	-	-	-	79.0	-	-	-	-
240	67.1	-	-	-	-	-	77.8	-	-	-	-
250	70.7	-	-	-	-	-	-	-	-	-	-
260	70.6	-	-	-	-	-	-	-	-	-	-
270	69.8	-	-	-	-	-	-	-	-	-	-
280	69.3	-	-	-	-	-	-	-	-	-	-
290	68.9	-	-	-	-	-	-	-	-	-	-
300	68.7	-	-	-	-	-	-	-	-	-	-
$E10^{(a)}$ %	77.5	73	84	83	81	91	95	86	85	96.5	80.5
Carbon Rate g/min.	-	7.23	7.70	9.76	4.58	-	4.2	3.98	4.69	4.25	-

(a) E10 = Sulphur removal efficiency at 10% CaO reacted

TABLE C-2

SHEET 3 of 3

SULPHUR REMOVAL IN FRESH BED BATCH REACTOR TESTS

% FUEL SULPHUR REMOVED

1-A Unit

Run. No: Time (min)	110	111	112	113	114	115	118	119	120	121
15	85.6	-	100	100	100	100	84.5	86.4	95.3	100
30	79.7	82.6	88.2	95.5	100	100	83.5	84.7	91.2	100
45	78.6	83.8	85.5	91.5	94.3	100	81.9	84.3	90.7	100
60	75.7	78.8	85.0	91.1	89.4	100	80.0	83.3	90.3	100
75	75.2	78.8	84.7	87.5	86.2	100	82.6	84.4	89.1	100
90	74.5	77.4	85.6	86.1	86.4	99.0	79.0	83.1	88.4	100
105	71.6	75.9	87.8	88.3	85.5	98.4	80.8	82.7	87.0	100
120	-	76.4	84.1	87.9	80.9	94.1	77.6	83.4	86.8	99.8
E10[a] %	73.0	77.0	85.0	87.5	86.2	98.8	79.5	83.5	88.0	100
Carbon Rate g/min.	7.57	2.54	1.11	1.67	-	0.55	6.89	6.61	4.17	0.59

(a) E10 = Sulphur removal efficiency at 10% CaO reacted

TABLE C-3

SHEET 1 of 3

SULPHUR AND METALS PICK UP IN BATCH CYCLIC TESTS

(UK Fuel and Limestone)

1-A Unit

Run No.	Cycle	% Fuel Sulphur Removed 15 min	20 min	35 min	40 min	60 min	Lime Metals Content(1) Sodium ppm	Vanadium ppm	Iron ppm	Bed Sulphur Content(2) After Regeneration % by wt.
33	1	-	-	-	-	93.8	160	280	640	1.07
34	2	-	91.0	-	94.7	100	180	500	680	1.42
35	3	-	92.5	-	94.1	93.8	170	660	670	1.75
36	4	-	93.7	-	95.1	91.4	200	840	780	2.21
37	5	-	91.7	-	-	90.2	200	1000	790	2.17
38	6	-	87.8	-	-	83.0	180	1090	910	1.77
39	7	-	-	-	-	-	160	1340	890	2.19
40	8	-	67.1	-	-	70.6	230	1340	890	1.99
41	9	-	-	-	-	-	150	1600	820	1.87
42	10	-	-	-	-	-	230	2000	1000	2.55
43	11	-	-	-	-	-	240	1900	820	2.79
44	12	-	-	-	82.5	73.8	270	2000	890	2.74
45	13	-	-	-	-	81.3	-	-	-	-
46	14	75.7	-	79.3	-	76.7	260	2400	880	2.62
47	15	83.5	-	82.6	-	80.2	260	2400	920	2.28
48	16	73.2	-	74.1	-	70.0	230	2600	1000	2.43
49	17	-	-	74.1	-	76.9	220	2600	1200	2.42
50	18	77.9	-	78.1	-	72.9	340	3200	1300	2.53
51	19	70.8	-	75.6	-	70.9	300	3100	1100	2.52
52	20	69.7	-	73.0	-	72.0	280	3100	1000	2.50
53	21	69.1	-	72.5	-	72.2	270	3100	1100	2.29

(1) Measured after each regeneration

(2) All sulphur present as sulphate

TABLE C-3

SHEET 2 of 3

SULPHUR AND METALS PICK-UP IN BATCH CYCLIC TESTS
(UK Fuel and US Limestone 1691)

1-A Unit

Run	Cycle	% Fuel	Sulphur	Removed	Lime	Metals Content*		Bed Sulphur	Content (% by wt)	
No.		15 min	35 min	60 min	Sodium ppm	Vanadium ppm	Iron ppm	Total S (60 min)	Total S (A.R.)	Sulphate (A.R.)
54	1	-	98.8	91.4	340	450	4600	3.02	2.01	1.98
55	2	89.3	87.1	89.5	320	650	4700	4.63	2.34	2.24
56	3	81.6	83.3	83.6	300	780	4500	5.44	3.16	2.31
57	4	-	84.7	80.3	180	760	4300	4.10	2.61	2.03
58	5	-	87.5	85.1	250	1050	4800	5.76	2.47	2.39
59	6	78.8	83.7	84.5	260	1130	5000	4.60	2.32	2.30
60	7	87.1	91.1	85.8	260	1380	4900	4.66	2.31	2.29
61	8	-	-	79.4	270	2130	4800	4.00	2.45	2.43
62	9	56.9	74.3	73.9	260	1820	4700	4.47	2.50	2.45
63	10	57.4	73.7	75.1	320	1800	4850	4.62	2.47	2.45
64	11	78.4	73.4	75.6	320	1790	4840	4.50	2.42	2.42
65	12	66.0	68.8	70.3	350	2030	4620	4.73	2.41	2.39
66	13	65.4	74.2	74.4	340	1900	4610	4.65	2.24	2.21
67	14	79.2	79.3	69.2	350	2310	5010	3.68	2.15	2.15
68	15	73.1	71.3	71.5	360	2360	4780	4.32	2.33	2.33
69	16	77.2	73.7	71.3	390	2450	4740	4.47	2.29	2.28
70	17	-	72.3	69.8	370	2030	4850	4.49	2.36	2.14
71	18	73.7	-	73.4	350	2230	4660	4.26	2.20	2.03
72	19	76.9	-	74.2	390	2540	4650	4.16	2.70	2.12

A.R. After Regeneration
* Measured after each regeneration

TABLE C-3

SULPHUR AND METALS PICK-UP IN BATCH CYCLIC TESTS
(UK Fuel and US Limestone 1690)

1-A Unit

Run	Cycle	% Fuel Sulphur Removed			Lime Metals Content*			Bed Sulphur Content (% by wt)		
No.		30 min	45 min	60 min	Sodium ppm	Vanadium ppm	Iron ppm	Total S (60 min)	Total S (A.R.)	Sulph (A.R.)
78	1	98.8	99.2	93.5	500	210	1450	3.34	1.58	1.58
79	2	92.1	89.3	85.8	460	360	1240	4.11	2.41	2.26
80	3	83.5	83.1	78.7	420	400	2590	4.37	1.81	1.81
81	4	82.9	82.2	79.9	480	430	1430	3.96	1.74	1.71
82	5	90.8	82.7	81.8	310	570	1350	3.83	1.89	1.83
83	6	88.1	86.1	84.8	290	880	5060	3.71	1.93	1.93
84	7	85.9	84.1	80.3	250	760	5050	4.00	1.56	1.56
85	8	87.7	85.7	82.0	290	730	5120	3.72	1.38	1.38
86	9	86.3	82.1	79.1	260	900	5170	3.57	1.43	1.43
87	10	86.0	82.1	77.1	230	1030	5040	-	1.37	1.35
88	11	-	79.2	76.7	260	1280	5100	3.70	1.85	1.84
89	12	78.8	77.7	74.1	230	1390	5070	3.87	2.15	1.97
90	13	77.0	72.8	70.8	260	1340	4780	3.86	1.94	1.84
91	14	68.0	-	65.8	240	1620	4950	3.89	1.95	1.82
92	15	-	-	-	250	1570	4790	3.95	1.81	1.72
93	16	65.6	66.8	64.4	220	2000	4900	4.04	2.29	2.01
94	17	75.9	78.8	73.0	260	1870	4700	4.13	2.29	2.14
95	18	72.7	64.3	64.1	260	2060	4740	4.30	2.13	2.04

A.R. After Regeneration

* Measured after each regeneration

TABLE C-4

CHANGE IN PARTICLE SIZE OF LIME DURING CYCLIC TEST

(UK Fuel and Limestone)

1-A Unit

Particle Size Range Microns	% of Bed in Size Range		
		Cycled Regenerated Lime	
	Uncalcined Stone	13 Cycles	20 Cycles
1200-1400	35.6	5.3	3.0
850-1200	57.9	53.3	49.0
600-850	5.8	28.7	32.5
300-600	0.0	10.9	13.8
< 300	0.7	1.8	2.7

TABLE C-5

CHANGE IN PARTICLE SIZE DISTRIBUTION DURING CYCLIC TEST

(US Limestone 1691)

1-A Unit

Particle Size Range Microns	% by wt. Bed in Size Range				
			Cycled Regenerated Lime		
	Uncalcined Stone	Calcined Stone	3 Cycles	13 Cycles	19 Cycles
1200-1400	30.5	20.9	30.7	14.4	23.0
850-1200	59.2	57.7	53.2	54.1	50.9
600-850	10.1	17.6	14.9	24.3	23.3
300-600	0	2.3	1.0	5.7	2.5
< 300	0.2	1.4	0.2	1.5	0.3

TABLE C-6

CHANGE IN PARTICLE SIZE DISTRIBUTION DURING CYCLIC TESTS

(US Limestone 1690)

1-A Unit

	% by wt. Bed in Size Range				
		Cycled Regenerated Lime			
Particle Size Range Microns	Uncalcined Stone	1 Cycle	5 Cycles	10 Cycles	18 Cycles
1200-1400	26.5	29.0	13.8	14.1	21.7
850-1200	64.1	55.4	49.8	49.9	48.2
600-850	8.9	13.8	24.3	24.3	20.4
300-600	0.1	1.2	7.0	8.3	5.9
< 300	0.4	0.6	5.1	3.4	3.8

TABLE C-7

COMPOSITION OF GASIFIER PRODUCT

DRY GAS BASIS

1-A Unit

Component	Run No:	Composition of Dry Gas, % by Vol.		
		53	65	86
Oxygen		0	0	0
Nitrogen		64.0	57.0	62.0
Carbon Monoxide		10.4	10.3	8.3
Carbon Dioxide		10.1	9.7	10.9
Hydrogen		4.2	10.0	7.4
Methane		4.4	6.5	5.1
C_2		5.4	5.8	5.4
C_3		0.3	0.5	0.8
C_4		0.2	0.2	0.1

APPENDIX D

BATCH REACTOR VARIABLE STUDY WITH FRESH LIME BEDS

The bulk of this appendix contains the data from the fresh bed tests which formed the initial part of the variable study using limestone BCR 1691 and the Amuay Fuel Oil. It also contains the results of commissioning tests on the 4A and 4B units.

As discussed in detail previously, the variable study was divided into two parts: fresh bed and cycle tests (4). Ideally, cycle tests should have been used throughout since we wished to relate the observed effect to continuous operation. However, they are very time consuming and a reasonable coverage of the variables in this manner would have been prohibitive in both time and cost. Consequently, the two part approach was adopted. The objective of the first part - the fresh bed test programme - was to determine the basic effects of the variables listed below and any interactions which may have occurred between them:

Air/fuel ratio
Bed Temperature
Superficial Gas Velocity
Bed Depth
Particle Size Range

It was also intended that these tests would allow some of the above variables to be left out of the second part of the variable study, the cycle test programme.

Previous studies had indicated the general region of conditions for reasonable CAFB operation. Therefore, the new variable study was concentrated in a fairly narrow range. The programme of fresh bed tests which were carried out is given in Table D-V. This was based on a sequential examination of the variables. The design was such that not only was detailed information on the basic effect of each variable obtained, but also indications of interactions which exist between variables. It is worth noting that a full factorial design had been considered but was rejected because our existing information on the behaviour of the variables was not sufficient to guarantee that the results from such a design would be meaningful. When, as in our case, it is possible to obtain results from individual runs before the entire experimental programme is completed, sequential experiment designs are much more efficient than factorial type designs.

The programme as listed differs slightly from the originally prepared programme (4), in that operation of the units showed, firstly, that higher air/fuel ratios than anticipated, were required to give bed temperatures in excess of 900°C and, secondly, that 870°C could not be achieved with 20% stoichiometric air at the lowest gas velocity (4 ft/sec). Consequently, all tests previously planned at 25% stoichiometric air and 1000°C were reprogrammed to 30% stoichiometric air and 950°C and the test at 20% stoichiometric air and 4 ft/sec gas velocity was abandoned. On-going analysis of results indicated that these alterations would not detract from the value of the tests.

The wide range lime particle size was investigated to determine the feasibility of using lower cost stone. Discussions with limestone manufacturers had indicated that the lowest cost size range of stone is the 1/8-down fraction which is nearly waste product in many operations. All previous tests had been carried out within the 600-1400 μ range.

In our data system each cycle in each unit has been given a run number which is suffixed by the unit number. Tables D-II to D-IV list fully the results from the commissioning tests and Tables D-VI to D-XIV the results from the fresh bed part of the variable study.

TABLE D 1

Test Conditions (Commissioning Tests 4A & 4B)

Run No.	Fuel	Lime-Stone	Limestone Particle Size Range μ	Fuel Rate (g/min)	Total Air Rate (l/min)	Superficial Gas Vel (ft/sec)	% Stoich. Air	Bed Depth (Static) (inches) Start	End	Average Bed Temp. Absorption °C	Max Bed Temp.Regeneration °C	Duration of Absorption (min)
1/4A	U.K.	U.K.	600-1400	155	440	4.0	26.0	15.5	14.2	855	1070	120
3 4A	U.K.	U.K.	600-1400	205	417	3.6	18.7	15.5	13.9	810	1080	120
21/4B	U.K.	U.K.	600-1400	182	413	3.6	20.8	15.5	14.4	805	1100	150
22/4B	U.K.	U.K.	600-1400	142	422	3.8	27.2	15.5	13.9	850	1050	175

TABLE D II

Sulphur Removal (Commissioning Tests)

Time(Min) :		15	30	45	60	75	90	105	120	135	150	165
% Fuel Sulphur Removed in :-	Run 1/4A	100	100	100	100	99.6	99.2	98.2	96.1	-	-	-
	Run 3/4A	70.9	74.4	66.6	66.2	59.4	67.1	56.8	57.9	-	-	-
	Run 21/4B	99.7	84.5	78.6	75.0	-	-	-	-	-	65.8	-
	Run 22/4B	100	100	100	100	100	100	99.1	98.4	96.6	95.0	91.8

TABLE D III

Regeneration Data (Unit 4A) (Commissioning Tests)

Time (min	Run 1/4A Exit Gas Composition (% by vol) CO2	CO	O2	SO2	Temperature °C	Run 3/4A Exit Gas Composition (% by vol) CO2	CO	O2	SO2	Temperature °C
1	19.2	0	0	-	890	-	-	0	-	860
1½	-	-	-	-	-	-	-	-	0	-
2	8.6	0	0	-	910	14.0	5.7	0	-	870
3	1.2	0	0	-	940	12.4	9.2	0	0	900
4	1.0	0	0	-	980	11.4	12.4	0	-	920
4½	-	-	-	-	-	-	-	-	0	-
5	0.7	0	0	-	1020	10.5	13.4	0	-	940
6	0.2	0	0	-	1025	10.0	15.0	0	0	960
7	0.1	0	0	-	1035	9.6	15.6	0	-	970
7½	-	-	-	-	-	-	-	-	0	-
8	0.1	0	0	-	1040	9.2	16.3	0	-	980
9	0	0	0	-	1050	9.6	16.3	0	0	1000
10	0	0	0	-	1060	9.6	16.0	0	-	1000
10½	-	-	-	-	-	-	-	-	0	-
11	0.2	0	0	-	1070	10.0	15.6	0	-	1010
12	2.5	0	9.2	-	1000	11.4	14.7	0	0	1020
13	4.4	0	10.8	-	950	14.6	12.4	0	-	1030
13½	-	-	-	-	-	-	-	-	0	-
14	3.6	0	12.8	-	910	10.5	5.7	0	-	1040
15	1.6	0	14.5	-	870	4.5	0.8	0	10.1	1060
16	-	-	-	-	-	-	0.2	0	-	1080
16½	-	-	-	-	-	-	-	-	13.6	-
17	-	-	-	-	-	3.5	0.2	0	-	1070
18	-	-	-	-	-	2.9	0.2	0	14.1	1070
19	-	-	-	-	-	2.5	0.2	0	-	1080
19½	-	-	-	-	-	-	-	-	14.7	-
20	-	-	-	-	-	2.2	0	0	-	1080
21	-	-	-	-	-	2.0	0	8.0	11.9	1080
22	-	-	-	-	-	5.1	0	11.8	-	1030
22½	-	-	-	-	-	-	-	-	0.41	-
23	-	-	-	-	-	7.0	0	13.1	-	1000
24	-	-	-	-	-	6.5	0	13.8	0.14	940
25	-	-	-	-	-	6.2	0	15.0	-	910
25½	-	-	-	-	-	-	-	-	0.04	-
26	-	-	-	-	-	5.2	0	16.0	-	870
27	-	-	-	-	-	3.2	0	16.8	-	850
28	-	-	-	-	-	2.[illegible]	0	17.[illegible]	-	820

TABLE D IV

Regeneration Data (Unit 4B) (Commissioning Tests)

Time (min)	Run 21/4B Exit Gas Composition (% by vol) CO2	CO	O2	SO2	Temperature °C	Run 22/4B Exit Gas Composition (% by vol) CO2	CO	O2	SO2	Temperature °C
1	13.4	8.4	0	-	820	-	-	-	-	880
1½	-	-	-	0	-	-	-	-	-	-
2	11.8	10.8	0	-	840	-	-	-	-	910
3	10.9	12.2	0	0	880	-	-	-	-	940
4	10.0	13.6	0	-	900	-	-	-	-	970
4½	-	-	-	0	-	-	-	-	-	-
5	9.6	14.5	0	-	920	-	-	-	-	1010
6	9.2	15.0	0	0	930	-	-	-	-	1020
7	8.8	15.0	0	-	950	-	-	-	-	1025
7½	-	-	-	0	-	-	-	-	-	-
8	8.4	15.5	0	-	960	-	-	-	-	-
9	8.4	15.5	0	0	970	-	-	-	-	1045
10	8.4	16.2	0	-	980	-	-	-	-	1055
10½	-	-	-	0	-	-	-	-	-	-
11	8.4	16.2	0	-	990	-	-	-	-	1060
12	8.4	16.2	0	0	995	-	-	-	-	1060
13	8.4	16.2	0	-	1000	-	-	-	-	985
13½	-	-	-	0	-	-	-	-	-	-
14	8.8	15.5	0	-	1010	-	-	-	-	950
15	9.2	15.0	0	0	1015	-	-	-	-	900
16	-	-	0	-	1020	-	-	-	-	-
16½	-	-	-	0	-	-	-	-	-	-
17	11.4	12.7	0	-	1025	-	-	-	-	-
18	12.9	10.0	0	0	1030	-	-	-	-	-
19	16.0	3.5	0	-	1040	-	-	-	-	-
19½	-	-	-	9.7	-	-	-	-	-	-
20	6.6	0.6	0	-	1050	-	-	-	-	-
21	4.2	0.2	0	12.2	1060	-	-	-	-	-
22	3.5	0.2	0	-	1080	-	-	-	-	-
22½	-	-	-	13.2	-	-	-	-	-	-
23	3.2	0.2	0	-	1080	-	-	-	-	-
24	3.0	0.2	0	13.4	1085	-	-	-	-	-
25	2.9	0	0	-	1090	-	-	-	-	-
25½	-	-	-	13.7	-	-	-	-	-	-
26	2.5	0	0	-	1095	-	-	-	-	-
27	2.2	0	0	13.0	1100	-	-	-	-	-
28	7.5	0	0	-	1100	-	-	-	-	-
28½	-	-	-	1.2	-	-	-	-	-	-
29	5.1	0	9.0	-	1090	-	-	-	-	-
30	6.0	0	12.1	0.38	1035	-	-	-	-	-
31	5.4	0	13.2	-	980	-	-	-	-	-
31½	-	-	-	-	-	-	-	-	-	-
32	4.8	0	14.0	-	960	-	-	-	-	-
33	4.2	0	14.9	-	920	-	-	-	-	-
34	3.5	0	16.2	-	890	-	-	-	-	-

TABLE D V

Fresh Bed Test Programme

Test	% Stoich Air	Temp °C	Gas Vel. ft/sec.	Reagent Particle Size (microns)	Static Bed Depth (in)
1	30	870	6	300 - 3000	15.5
2	25	870	6	"	15.5
3	20	870	6	"	15.5
4	15	870	6	"	15.5
5	25	800	6	"	15.5
6	30	950	6	"	15.5
7	20	800	6	"	15.5
8	25	870	4	"	15.5
9	25	870	8	"	15.5
10	20	870	8	"	15.5
11	25	870	6	600 - 1400	15.5
12	25	870	6	1200 - 3000	15.5
13	20	870	6	600 - 1400	15.5
14	20	870	6	1200 - 3000	15.5
15	25	870	6	300 - 3000	10
16	25	870	6	"	20
17	20	870	6	"	10
18	20	870	6	"	20
19	30	950	4	"	15.5
20	30	950	8	"	15.5
21	30	950	6	600 - 1400	15.5
22	30	950	6	1200 - 3000	15.5
23	30	950	6	300 - 3000	10
24	30	950	6	300 - 3000	20
25	25	870	8	600 - 1400	15.5
26	25	870	8	1200 - 3000	15.5
27	25	870	8	300 - 3000	10
28	25	870	8	300 - 3000	20
29	25	870	6	1200 - 3000	10
30	25	870	6	1200 - 3000	20

TABLE L VI

Fresh Bed Test Conditions

Test	Run No	Fuel	Limestone	Limestone Particle Size Range (μ)	Fuel Rate (g/min)	Total Air Rate (l/min)	Superficial Gas Velocity (ft/sec)	% Stoich. Air	Bed Depth (Static) inches Start	Bed Depth (Static) inches End	Average Bed Temp.Absorp. °C	Maximum Bed Temp. (Regen.)	Duration of Absorp. (min)	Distributor
1	24/4B	Amuay (2.3%S)	US(BCR1691)	300-3175	192	657	6.1	31.4	15.5	12.8	870	1090	180	Normal *
2	27/4B	"	"	"	247	617	5.6	23.3	15.5	13.0	850	1080	150	Normal *
3	6/4A	"	"	"	291	625	5.6	19.7	15.5	9.8	845	1130	120	High p.drop
4	28/4B	"	"	"	370	600	5.1	14.8	15.5	13.2	780	1055	[illegible]	Normal *
5	8/4A	"	"	"	221	631	5.5	26.1	15.5	11.4	800	1060	150	"
5A	10/4A	"	"	"	214	620	5.4	26.6	15.5	11.5	800	1060	150	"
6	27/4A	"	"	"	176	643	6.3	33.6	15.5	13.2	950	-	150	"
6A	29/4B	"	"	"	188	636	6.1	31.0	15.5	11.5	915	1090	150	"
7	26/4B	"	"	"	310	614	5.5	18.2	15.5	10.1	800	1095	120	"
8	7/4A	"	"	"	152	471	4.3	28.5	15.5	12.6	850	1060	180	High p.drop
9	12/4A	"	"	"	263	846	7.9	29.5	15.5	11.5	880	1060	150	Normal *
10	26/4A	"	"	"	359	810	7.3	20.6	15.5	11.9	845	1100	100	"
11	19/4A	"	"	600-1400	239	649	6.0	24.9	15.5	12.5	875	1100	150	"
12	45/4B	"	"	1200-3175	233	613	5.6	24.2	15.5	13.9	863	-	150	"
13	48/4B	"	"	600-1400	288	658	5.9	20.9	15.5	12.7	845	1080	120	"
14	37/4B	"	"	1200-3175	276	646	6.0	21.5	15.5	12.8	870	1100	120	"
15	24/4A	"	"	300-3175	220	634	5.8	26.4	10.0	7.2	870	1070	100	"
16	13/4A	"	"	"	230	631	5.8	25.1	20.0	12.3	870	1120	200	"
17	15/4A	"	"	"	285	646	5.9	20.8	10.0	8.7	870	1080	80	"
18	29/4A	"	"	"	268	633	5.8	21.5	20.0	18.8	865	1095	180	"
19	[illegible]8/4A	"	"	"	105	428	4.3	37.3	15.5	12.7	950	1020	180	"
20	32/4A	"	"	"	205	827	8.2	37.0	15.5	11.9	950	1045	120	"
21	20/4A	"	"	600-1400	163	606	6.0	34.1	15.5	11.2	945	1110	160	"
22	39/4B	"	"	1200-3175	171	598	5.9	31.9	15.5	14.2	950	1050	160	"
23	44/4B	"	"	600-1400	163	628	6.1	35.4	10.0	8.8	945	[illegible]	100	"
24	18/4A	"	"	300-3175	151	634	6.3	38.5	20.0	14.0	960	1100	220	"
25	21/4A	"	"	600-1400	320	855	7.7	24.5	15.5	11.9	850	1100	120	"
26	22/4A	"	"	1200-3175	323	825	7.7	23.4	15.5	12.5	890	1110	120	"
27	42/4B	"	"	300-3175	315	835	7.8	24.5	10.0	8.4	885	1000	[illegible]	"
28	30/4A	"	"	"	328	843	7.7	23.6	20.0	13.2	855	1100	[illegible]	"
29	38/4B	"	"	1200-3175	236	624	6.0	24.3	10.0	9.8	890	1080	100	
30	23/4A	"	"	"	213	631	5.9	27.2	20.0	17.6	875	1140	200	"

* Large pieces of refractory varying in size between ½" and ¼" were placed on the distributor.

Table B VII

Sulphur Removal (Fresh Bed Tests)

% Fuel Sulphur Removed

Test Time (min)	1	2	3	4	5	5A	6	6A	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
15	100	-	96.2	78.1	99.4	99.1	97.5	100	99.1	99.8	-	100	99.2	95.6	100	100	100	99.6	90.3	99.5	100	100	91.7	98.3	100	99.5	99.4	-	99.6	97.7	67.0	100
30	100	100	96.3	71.5	99.4	-	94.0	99.3	93.5	99.8	90.4	98.4	99.2	89.1	98.5	66.0	98.3	99.6	93.3	98.9	99.1	100	90.3	58.8	36.4	99.4	98.6	88.1	97.6	97.8	61.6	100
45	100	99.5	-	61.0	99.6	-	99.0	99.3	91.7	98.9	72.0	95.5	96.7	76.0	98.5	45.5	80.1	96.6	77.4	98.7	98.1	99.1	84.1	53.0	22.3	99.2	67.8	67.0	97.5	97.7	55.0	96.3
60	100	99.3	98.1	-	98.8	99.2	81.8	76.1	87.2	99.5	65.1	90.5	96.7	62.6	98.5	99.3	70.8	99.5	81.4	97.3	-	77.9	49.5	53.6	10.6	99.6	58.5	55.8	41.9	96.4	52.6	91.8
75	100	99.2	98.7	48.5	98.5	99.3	30.5	73.[illegible]	83.7	99.5	-	72.6	94.6	48.0	97.7	99.4	50.0	99.5	69.6	97.2	52.0	56.9	38.3	38.0	6.8	99.2	62.0	45.3	22.8	95.1	38.3	85.2
90	100	99.1	93.9	-	99.1	99.3	18.5	64.[illegible]	63.0	100	66.3	68.4	99.1	41.7	96.2	33.3	40.6	97.6	-	97.1	44.4	19.5	28.2	40.5	17.7	98.8	59.8	48.9	-	96.0	-	76.5
105	-	98.5	83.9	-	99.0	[illegible]	23.3	63.4	63.0	99.6	59.8	-	86.8	36.2	-	38.2	-	79.6	-	95.5	29.3	19.1	25.5	26.7	-	98.6	62.8	46.2	-	83.0	-	76.4
120	98.7	98.5	75.5	-	[illegible]	[illegible]	[illegible]	53.5	[illegible]	[illegible]	[illegible]	-	22.9	-	-	28.4	-	71.3	-	[illegible]	27.3	24.4	20.9	30.5	-	97.3	57.0	40.6	-	63.8	-	73.9
135	96.5	-	-	-	-	[illegible]	-	47.0	-	[illegible]	59.8	-	74.7	-	-	-	-	64.4	-	92.1	6.2	-	-	26.3	-	70.2	-	-	-	57.1	-	69.9
150	-	95.5	-	-	-	-	-	47.0	-	70.6	-	-	-	-	-	-	-	58.0	-	91.5	10.8	-	-	22.8		63.6	-	-	-	55.3	-	68.4
165	-	-	-	-	-	-	-	-	-	75.0	-	-	-	-	-	-	-	51.0	-	92.3	14.5	-	-	-	-	68.1	-	-	-	-	-	65.2
180	-	-	-	-	-	-	-	-	-	75.0	-	-	-	-	-	-	-	36.3	-	40.5	-	-	-	-	-	-	-	-	-	-	-	57.9
195	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	27.3	-	-	-	-	-	-	-	-	-	-	-	-	-	62.6
210	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
225	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-										

TABLE VIII

Regeneration CO$_2$ Concentrations (Fresh Bed Tests)

CO$_2$ Concentration (% by vol)

Test	1	2	3	4	5	6	6A	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
Time (min)																															
1	1.5	3.5	11.8	-	-	-	2.2	-	14.6	5.7	14.5	8.8	14.6	11.9	11.9	7.2	5.1	7.0	8.8	7.8	7.7	10.0	2.0	6.7	3.8	3.5	8.6	4.9	9.4	2.0	11.9
2	0.7	2.5	10.9	-	14.3	-	0.7	10.9	5.4	3.2	14.6	3.8	14.7	8.4	2.9	1.5	1.8	3.9	3.8	7.6	9.1	9.1	7.0	8.5	1.4	2.8	6.0	9.6	7.2	2.4	11.9
3	0.7	2.2	7.0	7.5	5.1	-	4.2	8.0	3.5	5.8	3.3	3.6	4.8	3.8	2.0	0.7	1.1	2.9	3.6	9.7	9.0	4.8	10.3	7.8	3.8	2.0	3.7	9.3	4.8	10.5	10.2
4	2.2	1.8	8.0	-	3.5	-	9.2	6.3	3.5	8.4	2.4	3.1	2.0	2.4	2.0	3.5	1.1	7.6	3.4	9.0	8.7	7.0	9.2	6.0	7.0	2.0	5.5	7.1	2.8	10.5	3.6
5	7.0	1.8	6.0	-	2.9	-	9.2	5.8	3.5	7.1	2.8	2.7	3.0	2.0	3.8	6.6	1.1	7.6	3.4	7.6	4.8	8.8	7.7	3.5	7.4	7.0	10.0	5.8	5.4	8.0	2.6
6	8.8	3.2	3.5	7.5	2.2	-	8.0	10.0	2.7	4.9	8.1	4.2	9.7	7.0	10.9	5.0	3.5	5.1	3.9	7.0	3.1	6.2	7.0	2.0	5.5	8.2	8.2	3.3	10.2	7.4	2.2
7	8.0	9.6	2.9	-	3.0	-	6.6	12.8	2.5	3.7	4.8	10.0	11.0	10.5	10.1	3.5	9.2	3.5	6.1	6.4	2.4	4.8	6.0	-	4.3	4.6	5.3	2.8	10.2	5.7	2.1
8	7.0	10.0	2.5	5.0	8.4	-	5.7	7.6	2.2	2.5	3.8	7.4	11.3	8.5	7.3	1.8	8.8	1.8	10.2	4.8	1.2	3.6	5.8	-	3.1	3.9	3.8	2.0	8.8	5.1	4.2
9	5.7	7.5	2.5	-	8.8	-	5.1	3.5	8.8	1.1	3.1	5.4	10.8	6.8	6.7	-	7.5	-	8.8	4.0	-	2.0	4.8	-	2.8	2.9	2.9	-	5.6	4.5	10.0
10	4.2	6.6	7.3	-	7.1	-	4.3	2.9	9.2	-	2.0	5.0	7.7	5.4	5.8	-	5.8	-	6.2	2.0	-	-	4.2	-	2.0	1.9	2.0	-	4.8	3.5	10.2
11	3.2	5.1	7.3	3.2	5.4	-	3.5	2.2	7.3	-	-	4.0	6.1	4.1	4.8	-	4.8	-	5.5	-	-	-	3.1	-	-	-	-	-	3.4	-	7.5
12	2.2	3.6	4.8	-	3.5	-	2.5	1.8	5.7	-	-	2.7	4.8	3.1	3.1	1	3.9	-	4.8	-	-	-	2.1	-	-	-	-	-	2.8	-	6.0
13	-	3.2	3.5	5.8	1.8	-	1.8	8.4	4.5	-	-	-	3.9	-	2.0	-	3.2	-	4.2	-	-	-	-	-	-	-	-	-	2.2	-	4.8
14	-	1.8	3.2	-	-	-	-	9.2	3.2	-	-	-	2.6	-	-	-	2.4	-	3.8	-	-	-	-	-	-	-	-	-	-	-	4.0
15	-	-	2.5	4.3	-	-	-	7.0	-	-	-	-	-	-	-	-	1.5	-	3.3	-	-	-	-	-	-	-	-	-	-	-	3.3
16	-	-	1.8	4.0	-	-	-	5.4	-	-	-	-	-	-	-	-	1.1	-	3.1	-	-	-	-	-	-	-	-	-	-	-	2.2
17	-	-	1.8	4.5	-	-	-	5.1	-	-	-	-	-	-	-	-	-	-	2.9	-	-	-	-	-	-	-	-	-	-	-	-
18	-	-	-	5.1	-	-	-	3.9	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
19	-	-	-	6.0	-	-	-	3.2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
20	-	-	-	6.0	-	-	-	2.5	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
21	-	-	-	6.0	-	-	-	1.8	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
22	-	-	-	6.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE D IX

Regeneration CO Concentrations (Fresh Bed Tests)

CO Concentration (% by vol)

Test Time (min)	1	2	3	4	5	6	6A	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
1	2.5	0	5.9	-	-	0	0	-	1.1	2.8	2.8	4.1	3.5	2.8	1.2	4.5	2.5	1.6	5.2	4.7	-	3.8	0	3.8	1.6	0.6	1.6	0.8	3.3	2.5	7.0
2	0	0	10.3	-	4.1	0	0	10.0	0	0	1.0	0.6	2.5	0	0	0	0.3	0.3	0.3	0	-	0.6	0	0	0.3	0	0.8	0	0.6	0	8.0
3	0	0	15.7	2.8	0	0	0	16.1	0	0	0.3	0.3	0	0	0	0	0	0	0	0	-	0	0	0	0	0	0.5	0	0	0	1.6
4	0	0	13.0	-	0	0	0	19.2	0	0	0	0.3	0	0	0	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0
5	0	0	0	-	0	0	0	20.0	0	0	0	0.3	0	0	0	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0
6	0	0	0	9.2	0	0	0	20.0	0	0	0	0	0	0	0	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0
7	0	0	0	-	0	0	0	13.6	0	0	0	0	0	0	0	0	0	0	0	0	-	0	0	-	0	0	0	0	0	0	0
8	0	0	0	16.2	0	0	0	1.6	0	0	0	0	0	0	0	0	0	0	0	0	-	0	0	-	0	0	0	0	0	0	0
9	0	0	0	-	0	0	0	0	0	0	0	0	0	0	0	-	0	-	0	0	-	0	0	-	0	0	0	-	0	0	0
10	0	0	0	-	0	-	0	0	0	-	0	0	0	0	0	-	0	-	0	0	-	-	0	-	0	0	0	-	0	0	0
11	0	0	0	18.1	0	-	0	0	0	-	-	0	0	0	0	-	0	-	0	-	-	-	0	-	-	-	-	-	0	-	0
12	0	0	0	-	0	-	0	0	0	-	-	0	0	0	0	-	0	-	0	-	-	-	0	-	-	-	-	-	0	-	0
13	-	0	0	8.4	0	-	0	0	0	-	-	-	0	-	0	-	0	-	0	-	-	-	-	-	-	-	-	-	0	-	0
14	-	0	0	-	-	-	-	0	0	-	-	-	0	-	0	-	0	-	0	-	-	-	-	-	-	-	-	-	-	-	0
15	-	-	0	3.2	-	-	-	0	-	-	-	-	-	-	-	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	0
16	-	-	0	1.9	-	-	-	0	-	-	-	-	-	-	-	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	0
17	-	-	0	1.4	-	-	-	0	-	-	-	-	-	-	-	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-
18	-	-	-	0.8	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
19	-	-	-	0.2	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
20	-	-	-	0.2	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
2	-	-	-	0.2	-	-	-	0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
22	-	-	-	0.2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE D X

Regeneration O_2 Concentrations (Fresh Bed Tests)

O_2 Concentration (% by vol)

Test	1	2	3	4	5	6	6A	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
Time (min)																															
1	0	0	0	0	0	-	0	0	0	0	0	0	0	0	0	0	0	0	0	0	-	0	0	0	0	0	0	1.0	0	0	0
2	0	0	0	0	0	-	0	0	0	2.0	0	0	0	0	0	0	0	0	0	3.0	-	0	4.0	8.0	0	0	0	6.0	0	0	0
3	0	0	0	0	0	-	2.0	0	0	3.6	0	0	0	0	0	3.0	0	0	0	5.6	8.0	0	7.0	11.0	1.5	0	0	9.0	0	5.0	0
4	0	0	0	0	0	-	5.0	0	0	8.0	0	0	0	0	0	6.0	0	5.0	0	9.0	8.5	1.0	9.2	14.0	5	0	1.8	11.0	0	8.0	0
5	0	0	0	0	0	-	8.0	0	0	11.0	2.0	0	6.0	0.5	1.5	10.1	0	9.0	0	11.2	13.0	6.0	11.0	16.0	10.0	1.5	7.6	13.8	2.0	11.1	0
6	0	0	0	0	0	-	12.0	0	0	14.0	9.0	1.0	4.0	3.0	6.5	13.2	0	12.4	0	13.0	15.0	7.8	12.0	17.0	12.0	8.0	10.0	15.0	4.0	12.2	0
7	8.8	0	0	0	4.0	-	14.0	0	0	15.6	13.2	6.0	7.1	7.0	8.0	15.0	0	15.0	0.8	13.4	16.2	12.0	13.6	-	14.1	12.0	12.8	16.0	8.0	13.9	0
8	11.1	7.8	0	0	8.0	-	13.0	0.5	0	17.0	15.0	10.0	7.9	9.0	11.5	17.0	5.6	16.0	1.6	15.0	17.0	15.0	13.8	-	15.1	15.0	15.0	17.0	11.2	14.4	1.0
9	12.0	11.8	0	0	10.1	-	14.0	0.5	6.8	18.2	16.2	13.0	9.0	12.5	13.0	-	5.8	-	5.0	16.0	-	16.0	15.0	-	16.2	16.5	16.0	-	13.0	15.2	5.0
10	14.1	14.0	4.7	0	12.3	-	15.0	0	8.0	-	18.0	14.8	12.2	14.0	14.0	-	5.8	-	8.0	18.0	-	-	15.4	-	16.9	18.0	17.6	-	14.6	16.5	6.4
11	14.9	14.9	8.8	0	13.9	-	15.0	0	11.0	-	-	16.0	13.1	16.0	15.5	-	11.0	-	11.0	-	-	-	16.2	-	-	-	-	-	16.0	-	8.6
12	15.8	16.1	11.5	0	16.0	-	16.0	1.5	13.0	-	-	16.4	15.0	17.0	16.5	-	12.0	-	13.1	-	-	-	17.0	-	-	-	-	-	16.5	-	11.0
13	-	17.1	13.8	0	-	-	19.0	7.0	15.0	-	-	-	15.8	-	17.8	-	13.0	-	14.2	-	-	-	-	-	-	-	-	-	17.0	-	13.6
14	-	17.9	15.0	0	-	-	-	9.0	15.3	-	-	-	17.8	-	-	-	15.0	-	15.0	-	-	-	-	-	-	-	-	-	-	-	14.6
15	-	-	15.5	0	-	-	-	14.0	-	-	-	-	-	-	-	-	[illegible]	-	15.6	-	-	-	-	-	-	-	-	-	-	-	15.2
16	-	-	16.0	0	-	-	-	14.0	-	-	-	-	-	-	-	-	16.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-
17	-	-	16.0	0	-	-	-	15.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
18	-	-	-	0	-	-	-	16.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
19	-	-	-	1.9	-	-	-	16.5	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
20	-	-	-	3.1	-	-	-	17.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
21	-	-	-	4.8	-	-	-	18.0	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
22	-	-	-	5.1	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

TABLE D XI

Regeneration SO_2 Concentrations (Fresh Bed Tests)

SO_2 Concentration (% by vol)

Test Time (min)	1	2	3	4	5	6	6A	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	23	24	25	26	27	28	29	30
1½	8.9	2.3	0	-	0	2.0	2.4	0	0.8	-	-	2.4	-	-	-	3.6	0	0	-	-	-	-	-	-	-	-	-	-	9.4	-
3	7.4	9.8	0	-	0	0.2	0.6	0	10.3	-	-	16.3	-	-	-	11.5	10.3	6.2	-	-	-	-	-	-	-	-	-	-	-	-
4½	1.5	10.6	0	-	9.8	0.1	0.2	0	14.8	-	-	16.1	-	-	-	0.7	2.4	-	-	-	-	-	-	-	-	-	-	-	0.1	-
6	0.3	3.2	12.4	-	12.2	-	-	0	15.2	-	-	3.4	-	-	-	0.1	0.5	0.8	-		-	-	-	-	-	-	-	-	-	-
7½	0.2	0.6	13.7	-	2.0	-	-	0.4	13.0	-	-	0.5	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
9	0.1	0.2	12.0	-	0.5	-	-	11.8	0.6	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
10½	-	-	1.0	-	-	-	-	13.2	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
12	-	-	0.1	-	-	-	-	1.9	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

Table D XII

Regeneration Temperatures ([illegible] Bed Tests)

Temperature °C

Test Time (min)	1	2	3	4	5	6	6A	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
1	930	-	850	750	-	960	980	800	860	900	875	890	-	900	900	910	875	870	975	975	980	990	985	960	970	900	900	920	920	960	890
2	980	900	915	760	860	930	1020	840	900	980	930	930	-	950	940	940	980	970	930	1020	1070	1020	1030	990	1000	950	950	1000	1000	1050	920
3	1040	950	940	790	930	-	1060	900	915	1030	1010	990	-	1000	990	1030	1030	1000	970	1005	1045	1070	1050	940	1056	1000	1030	955	1030	1085	950
4	1090	930	990	815	980	860	1090	940	955	980	1040	1030	-	1040	1050	1060	1050	1080	1010	955	970	1110	1000	890	1100	1040	1100	870	1060	1030	-
5	1090	1010	1030	840	1000	830	1000	955	985	945	1090	1050	-	1060	1100	1070	1080	1040	1050	920	910	1090	950	840	1050	1100	1110	830	1100	990	990
6	1090	1030	1060	870	1020	800	940	970	995	880	1100	1080	-	1080	1090	1010	1120	980	1080	885	850	1020	900	-	1020	1020	1030	770	1080	900	1050
7	1030	1070	1080	-	1060	750	910	1000	1015	810	970	1100	-	1040	1030	930	1100	890	1095	850	815	990	880	-	950	930	990	750	1040	860	1080
8	960	1030	1090	910	1045	-	860	1035	1035	750	880	1040	-	960	910	865	1030	840	1040	820	790	925	835	-	910	850	910	730	950	810	1100
9	910	965	1110	930	1005	660	920	1040	1050	700	835	970	-	870	920	-	950	-	970	800	-	880	800	-	850	810	820	-	875	790	1120
10	840	930	1130	950	980	-	790	1060	1060	-	790	900	-	840	860	-	890	-	940	-	-	-	770	-	840	750	790	-	815	-	1140
11	820	860	1040	950	870	-	-	1060	990	-	-	840	-	820	815	-	850	-	890	-	-	-	750	-	-	-	-	-	760	-	1070
12	800	830	970	980	815	-		1070	950	-	-	820	-	780	770	-	820	-	840	-	-	-	720	-	-	-	-	-	730	-	1010
13	-	815	900	1000	-	-	-	1080	900	-	-	-	-	-	730	-	790	-	820	-	-	-	-	-	-	-	-	-	710	-	980
14	-	780	870	1020	-	-	-	1095	880	-	-	-	-	-	-	-	730	-	785	-	-	-	-	-	-	-	-	-	-	-	930
15	-	-	830	1030	-	-	-	1030	-	-	-	-	-	-	-	-	-	-	765	-	-	-	-	-	-	-	-	-	-	-	880
16	-	-	800	1050	-	-	-	960	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
17	-	-	740	1040	-	-	-	880	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
18	-	-	-	1015	-	-	-	850	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
19	-	-	-	940	-	-	-	800	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
20	-	-	-	910	-	-	-	780	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
21	-	-	-	880	-	-	-	[illegible]	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
22	-	-	-	835	-	-	-	725	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-

Table D XIII

Stone Carbon and Sulphur Contents (Fresh Bed Tests)

Test	Stone Carbon Content After Absorption (% by wt) *	Stone Sulphur Content After Regeneration (% by wt)
1	0.04	-
2	0.38	2.43
3	4.68	1.19
4	11.88	0.98
5	1.59	2.02
6	0.06	0.35
7	7.12	-
8	0.65	2.01
9	0	3.14
10	1.84	2.48
11	0.19	3.02
12	0.50	2.85
13	0.77	2.24
14	0.79	1.97
15	-	1.48
16	0	-
17	0.12	1.78
18		3.13
19	0.06	0.54
20	0.04	-
21	0.09	0.98
22	0	1.13
23	-	-
24	0.08	-
25	0.24	2.36
26	0.24	2.57
27	0.12	2.03
28	-	2.56
29	0.23	2.28
30	1.00	2.16

* Sample taken from top of bed

TABLE D XIV

Gasifier Gas Analysis (Fresh Bed Tests)

Test	Component Concentrations (% by vol) *										% Fuel C Oxidised	% Fuel $C > C_4$	% Fuel H Oxidised
	O_2	N_2	CO_2	CO	CH_4	C_2H_4	C_2H_6	C_3	C_4	H_2			
1	0	57.8	6.7	15.3	4.7	3.5	0.2	0.2	0.2	10.5	63.6	0	9.4
2	0	60.8	8.2	6.0	5.5	5.1	0.4	0.7	0.3	9.8	26.4	33.5	28.0
3	0	57.5	10.6	7.8	6.5	6.0	0.4	1.0	0.4	8.7	34.1	21.7	4.1
4	0	59.1	11.6	3.3	6.9	7.4	1.0	3.3	1.7	5.7	20.2	25.2	8.7
5	0	64.5	11.4	6.3	3.5	4.1	0.4	1.4	0.6	5.4	38.6	20.1	15.0
6	0	62.7	6.8	13.7	3.6	2.0	0.2	0.1	0.1	10.4	58.3	17.2	24.5
7	0	53.5	10.8	5.4	6.6	6.7	0.6	1.6	0.7	6.9	29.3	18.0	4.2
8	0	56.1	5.9	17.8	4.3	3.5	0.3	0.3	0.1	10.7	64.6	0	2.1
9	0	62.9	9.7	10.8	3.8	3.8	0.3	0.6	0.1	7.3	51.6	12.4	11.3
10	0	62.0	9.9	9.4	6.0	6.1	0.2	0.8	0.4	7.2	34.5	24.8	8.9
11	-	-	-	-	-	-	-	-	-	-	-	-	-
12	0	55.1	6.4	15.5	4.7	3.9	0.3	0.3	0.2	10.6	51.5	13.8	4.3
13	0	58.4	9.6	11.7	6.3	5.2	0.4	0.4	0.3	7.7	41.3	16.3	0.4
14	0	60.4	8.0	12.6	5.9	5.3	0.4	0.4	0.1	9.2	39.2	24.8	9.6
15	0	63.7	7.8	11.6	3.8	3.2	0.2	0.1	0.1	7.8	43.4	31.5	19.7
16	0	55.2	7.8	11.8	5.1	5.1	0.3	0.3	0.1	8.3	48.2	8.6	5.5
17	0	57.5	8.3	12.0	5.1	5.1	0.5	0.7	0.3	9.2	39.3	22.5	6.1
18	-	-	-	-	-	-	-	-	-	-	-	-	-
19	0	63.0	7.7	15.9	2.4	1.9	0.1	0.1	0.1	8.7	75.6	10.0	10.5
20	0	57.9	6.4	13.3	2.3	1.9	0.1	0.1	0.1	9.0	67.6	8.4	21.5
21	0	62.0	4.9	16.3	3.6	2.3	0.1	0.2	0.1	11.5	62.1	2.8	28.0
22	0	59.4	5.3	18.1	3.3	1.3	0.1	0.1	0.1	12.8	68.0	12.3	11.5
23	0	58.4	5.0	17.2	3.6	1.3	0.1	0.1	0.1	12.7	72.1	4.8	17.0
24	0	64.6	6.5	14.0	3.3	1.6	0.1	0.1	0.1	9.9	65.7	10.7	30.9
25	0	61.4	7.0	12.6	4.1	4.0	0.2	0.8	0.1	8.4	41.8	25.4	17.6
26	0	58.7	9.0	9.2	5.4	5.8	0.4	0.8	0.3	8.0	39.0	15.2	11.0
27	0	58.2	7.3	12.2	6.1	4.7	0.4	0.2	0.1	10.6	43.5	17.7	[illegible]
28	0	62.1	12.2	-	7.5	5.4	0.4	0.4	0.3	4.2	-	-	-
29	0	55.9	7.3	12.0	6.5	5.3	0.4	0.2	0.1	9.8	44.8	11.2	10.1
30	0	56.4	6.1	16.4	4.1	3.2	0.2	0.4	0.1	10.1	57.9	12.1	5.8

* Measured at ambient conditions

The data in Table D XIV were subjected to regression analysis and equations derived to describe the effects of operating variables on the conversion of the fuel oil to the various components. These equations which have been used to extrapolate the data are listed below.

a) % by wt C as CH_4 $= 1.35 \times \frac{T^{0.9} \times V^{0.1}}{e^{2.8} \times R^{0.4}}$

b) % by wt C as C_2H_4 $= 1.5 \times \frac{e^{25.8} \times V^{0.4}}{T^{3.4} \times R^{0.28}}$

c) % by wt C as C_2H_6 $= 1.25 \times \frac{e^{14.6}}{R^{0.9} \times T^{1.5} \times V^{0.4} \times D^{0.1}}$

d) % by wt C as C_3 $= \frac{e^{67.6} \times R^{0.5} \times V^{0.3} \times D^{0.3} \times S^{0.2}}{T^{10.5}}$

e) % by wt C as C_4 $= \frac{T^{2.2} \times D}{e^{6.0} \times R^{2.1} \times V^{0.9} \times S^{0.4}}$

f) % by wt $C > C_4$ $= \frac{T^{0.9} \times V^{5.1} \times S^{0.4}}{e \times R^{4.4} \times D^{0.2}}$

g) % by wt C oxidised $= 1.1 \times \frac{R^{0.9} \times T^{1.3} \times S^{0.1}}{e^{8.2} \times V^{0.3}}$

h) CO/CO_2 molar ratio $= 0.000291 \times e^{0.00976\,T}$

i) % by wt H as H_2 $= \frac{R^{0.9} \times T^{2.4} \times S^{0.1}}{e^{16} \times V^{0.3} \times D^{0.1}}$

j) % by wt H oxidised $= \frac{e^{35.5} \times R^{3.2} \times V^{2}}{T^{6.7} \times D^{0.7}}$

R = Air/Fuel Ratio

T = Bed Temperature

V = Gas Velocity

D = Bed Depth

S = Number Average Particle Size

APPENDIX E

BATCH REACTOR CYCLE TEST VARIABLE STUDY

This appendix contains the data from the cycle tests which formed the second part of the variable study using limestone BCR 1691 and the Amuay Fuel Oil. In this part, the following variables were examined.

Bed Replacement Rate
Gas Residence Time
Projected Sulphur Differential (P.S.D)
Particle Size Range

Our previous experience had shown that bed replacement rate had an important effect on bed deactivation during cycle tests (4) and, consequently, this variable was selected for investigation. The fresh bed tests showed that gas velocity and bed depth could be combined as residence time which had a marked effect on sulphur removal. It was considered that this variable warranted further examination by cycle tests. The third variable to be examined was projected sulphur differential. It was not originally planned to look at the effects of this variable since satisfactory sulphur removal was maintained in the fresh bed tests to bed sulphur levels far in excess of the 3 to 4% proposed for the cycle tests. However, as the cycle tests progressed it became apparent that this particular variable should also be investigated. The fourth variable: particle size range, fell into the same category as P.S.D.

On the basis of the results from the fresh bed tests, the target levels for the other variables were set as follows:-

Air/fuel ratio	25% of stoichiometric
Bed Temperature	870°C

Although the fresh bed tests had shown satisfactory sulphur removal down to 20% stoichiometric air, 25% was selected on the grounds of more convenient operation during cycle tests. Insufficient heat is liberated at the lower air/fuel ratio for 870°C to be readily maintained during cycle tests in the batch units.

The programme of tests was of sequential design with the objective being to outline operating conditions for minimum make-up rate and maximum gas velocity. Minimum make-up rate is desired to minimise operating cost and maximum gas velocity to minimise capital cost of a commercial gasifier.

The test results are summarised in Table E-I and listed fully in Tables E-II to E-XXVIII.

The limestone comparison test is listed first. This test was made during the programme to select between available U.S. limestones (4). In this test, BCR 1691 gave 72% lined out efficiency at 0.75 wt CaO/wt. sulphur. The lined out efficiency was identical to that obtained with the Denbighshire stone at the same make-up rate. At this stage, therefore, it looked as if the U.S. stone would behave similarly to the U.K. stone under cyclic conditions. This surmise was supported by results from the fresh bed tests. However, as the rest of the data in Table E-I show, appreciable differences were revealed as the cycle tests on BCR 1691 progressed.

The first test in the current series (T1C1 to C14) was carried out at a lime replacement rate of 2.4 wt CaO/wt. sulphur. This rate would have given a lined out efficiency of nearly 100% under the original test conditions with UK stone, and on the basis of the comparison test and the fresh bed studies, it was expected to give a similar high efficiency with BCR 1691. In fact, the efficiency lined out at only 61%, which is lower than that observed in the comparison test at lower make-up rate. A repeat test, T3C1 to C14, gave the same result.

These BCR 1691 stone cyclic tests differed from the original comparison test in a number of respects as shown below.

	Comparison Test	Recent Tests
Superficial Air Velocity, ft/sec.	4	6
Lime Particle Size, microns	600 - 1400	300 - 3200
Fuel Oil	U.K.	U.S. (Amuay)
Vanadium in Oil ppm	130	366
Sulphur in Oil wt.%	3.03	2.30
Unit	1-A	4A, 4B

The next series of cyclic tests was conducted to isolate each of these differences in turn in order to establish which, if any of these, was responsible for the lower efficiency.

Reducing superficial gas velocity to 4 ft/sec, in tests T1 C15 to C19 and T3 C15 to C23, increased lined out sulphur removal efficiency to 75%.

In tests T3 C27 to C31, make up with 600-1400 micron stone improved lined out efficiency to 68%. At first, this seems contrary to the fresh bed test results. However, the difference can be explained by the fact that the wider particle size range stone suffers greater loss during calcination than the 600-1400 micron material. Separate tests show an average loss of 52.5% on calcination of the 300-3200 micron stone compared with 42.7% of the 600-1400 micron size range. The effective makeup rate of the wide range stone was, therefore, lower than that of the narrow range stone at equal total addition rates. This effect did not operate in fresh bed tests, because in those tests, equal weights of precalcined lime were added to the reactor.

Test T4, cycles C1 to C10, was made with the U.K. fuel oil containing 3.03% sulphur and 130 ppm vanadium. Sulphur removal efficiency was only 62%, the same as obtained with the U.S. fuel. The cause of the difference, therefore, was not the fuel oil.

These tests indicated that the cause of lowered sulphur removal efficiency in the cyclic tests was a combination of increased gas velocity and greater lime losses during in-situ calcination of the wide particle size lime. No test of differences between experimental reactors was made, as earlier comparisons had shown no differences between the No.1 and No.4 units. It is unlikely that a difference would arise with a different limestone.

Although reverting to a narrow particle size stone and reducing gas velocity would improve sulphur removal efficiency, both of these changes would increase costs. Therefore, efforts have been directed at finding conditions which would allow use of the high velocity and wide size range while improving sulphur pick-up and reducing replacement requirements.

The tests discussed above were all conducted at a static lime bed depth of 15.5 inches. When lime replacement rate at this bed depth was increased to 3.09 wt CaO/wt. sulphur, the sulphur removal efficiency improved slightly, to 65% (T3 C32 to C41).

On further increasing the replacement rate to 5.08 wt. CaO/wt.S, in a test starting with a fresh bed, an unusual reversal in trend was noted. As usual in the first few cycles after starting with a fresh bed, the efficiency declined. It fell to below 80%, but then began rising again. The improved performance paralleled an increase in bed level as measured by bed pressure drop. The bed level rise was due to insufficient lime being removed between cycles to compensate the high replacement rate.

Level was allowed to rise to the equivalent of a 20 inch static bed while the test series continued. The efficiency lined out at 98% at this level and replacement rate.

Reducing replacement rate to 2.5 wt CaO/wt.S, while maintaining a 20 inch bed, gave a sulphur removal efficiency of 90% (T6 C1 to C17).

The improvement in performance with bed depth was not entirely due to greater residence time. The test at 15.5 inches with 4 ft/sec. superficial gas velocity actually had greater residence time than the 20 inch bed test at 6 ft/sec. but gave only 74% lined out efficiency. A contributing factor was thought to be the reduction in sulphur differential, the total sulphur pick-up per cycle, in the 20 inch bed tests. This reduction came about because cycle length was held constant while lime charge in the bed increased in going from 15.5 to 20 inches. In the 15.5 inch bed tests the PSD was 2.6 wt.%. In the 20 inch tests this projected differential was only 2.2%.

Subsequent tests were mainly directed at pursuing these trends further with the objective of maintaining high sulphur removal under lined out conditions at the lowest replacement rate.

Test 8, cycles 7 to 20, shows that an increase in static bed depth to 22.5 inches and a reduction in PSD to 2.0 wt.%. gave 97% sulphur removal at the significantly lower replacement rate of 2.5 wt. CaO per wt. sulphur. Attempts to pursue these trends even further (Test 8 cycles 45 to 61, Tests 10 and 11) were unsuccessful because satisfactory regeneration temperatures were not reached. The deeper beds increased the heat losses through the walls of the units. This heat loss is critical with the small batch units but is not a problem witn the much larger continuous unit.

Test 8, cycles 67 to 80, and to a greater extent, Test 9 separate the PSD and residence time effects. The separation was achieved by operating with a shallow bed to give a low residence time whilst reducing the length of the gasification state of each cycle to give a lower PSD. Test 9, which is similar in all respects to Test 1 with the exception of PSD being reduced from 3.0 wt% to 1.9%, gave a lined out sulphur removal efficiency of 86% as opposed to 61%. This indicates that changes in PSD are more effective than changes in residence time. Reductions in PSD are, however, limited by the amount of sulphur required on the stone to maintain adequate regeneration temperatures.

A base case, Test 12, with U.K. fuel and stone, was carried out at the end of the BCR 1691 cycle tests to ensure cycle test results from the OAP batch units could be safely correlated with those from previous units. The lined out efficiency of 89% when compared with the 92% expected from the relationship between make-up rate and lined out efficiency determined previously (Interim Report) indicates that the OAP batch units have behaved similarly to previous units.

TABLE E 1

Cycle Test Summary Table using BCR 1691

Test	Fuel	Limestone Particle Size Range (µ)	Absorption Temperature °C	% Stoich. Air	Superficial Gas Vel. (ft/sec)	Bed Depth (Static) (inches)	Residence Time$^{(a)}$ (sec)	Make-up Rate (WtCaO/ Wt.S)	Sulphur Differential (wt.%)	Lined Out Efficiency %
Comparison	UK (3.03%S)	600-1400	870	25	4	15.5	0.22	0.75	3.7	72
T1 (C1 to C14)	US (2.3%S)	300-3175	"	"	6	"	0.15	2.43	3.0	61
T1 (C15 to C19)	"	"	"	"	4	"	0.22	2.76	2.6	75
T3 (C1 to C14)	"	"	"	"	6	"	0.15	2.38	3.0	61
T3 (C15 to C23)	"	"	"	"	4	"	0.22	2.75	2.6	74
T3 (C24 to C31)	"	"	"	"	6	"	0.15	2.34 (600-1400µ)	3.1	68
T4 (C1 to C10)	UK (3.03%S)	"	"	"	6	"	0.15	2.24	3.2	62
T3 (C32 to C41)	US (2.3 %S)	"	"	"	6	"	0.15	3.09	3.1	65
T5 (C1 to C20)	"	"	"	"	6	20	0.21	5.08	2.2	98
T6 (C1 to C17)	"	"	"	"	6	"	0.21	2.50	2.2	90
T8 (C7 to C20)	"	"	"	"	6	22.5	0.24	2.50	2.0	97
T8 (C23 to C34)	"	"	"	"	6	"	0.24	1.70	2.2	79
T8 (C45 to C61)	"	"	"	30	6	25.0	0.28	2.69	1.7	77
T8 (C67 to C80)	"	"	"	25	6	15.5	0.15	2.16	2.2	71
T9	"	"	"	"	6	15.5	0.15	2.55	1.9	86
T10	"	"	"	35	4	22.5	0.36	2.68	1.5	89
T11	"	"	"	27.5	5	20	0.25	2.12	2.2	77
T12	UK (3.03%S)	UK(600-1400)	"	30	4	15.5	0.22	1.32	3.4	89

(a) Residence time in bed region above fuel injector.

TABLE E I1

Test Conditions (Cycle Test 1)

Run No.	Cycle No.	Fuel	Lime-Stone	Limestone Particle Size Range (μ)	Fuel Rate (g/min	Total Air Rate (1/min)	Superficial Gas Vel. (ft/sec)	% Stoich. Air	Bed Depth (Static) (inches) Start	End	Average Bed Temp. (Absorption) (°C)	Max. Bed Temp. (Regeneration) (°C)	Duration of Absorption (min.	Distri-butor	Make-up weight (limestone (g/cycle)
52/4B	1	#3(2.3%S)	BCR(1691)	300-3175	259	652	5.9	23.1	13.3	12.8	850	1000	45	Normal*	1200
53/4B	2	"	"	"	232	676	6.2	26.7	14.0	14.1	860	965	"	"	"
54/4B	3	"	"	"	238	632	5.7	24.4	16.7	15.1	835	1020	"	"	"
55/4B	4	"	"	"	219	670	6.1	28.1	18.3	16.9	865	1030	"	"	"
56/4B	5	"	"	"	221	646	5.9	26.8	19.3	18.0	860	1040	"	"	"
57/4B	6	"	"	"	221	675	6.2	28.0	20.0	17.7	870	1030	"	"	"
58/4B	7	"	"	"	240	683	6.3	26.1	20.7	17.3	865	1050	"	"	"
59/4B	8	"	"	"	246	672	6.2	25.1	20.0	17.9	865	1030	"	"	"
33/4A	9	"	"	"	240	629	5.7	24.0	15.5	-	845	-	"	"	"
34/4A	10	"	"	"	243	673	6.1	25.4	-	-	850	1020	"	"	"
35/4A	11	"	"	"	233	671	6.1	26.4	-	-	860	1030	"	"	"
36/4A	12	"	"	"	232	666	6.1	26.3	-	-	855	1040	"	"	"
37/4A	13	"	"	"	216	681	6.3	28.9	-	-	875	1040	"	"	"
38/4A	14	"	"	"	229	660	6.0	26.4	-	-	850	1020	39	"	"
60/4B	15	"	"	"	139	410	3.7	27.1	14.0	13.3	860	1000	68	"	"
61/4B	16	"	"	"	132	437	4.0	30.4	15.0	14.0	870	1035	"	"	"
62/4B	17	"	"	"	136	429	3.9	28.7	14.7	14.7	860	1020	"	"	"
63/4B	18	"	"	"	137	432	4.0	28.9	14.7	14.0	865	1060	"	"	"
64/4B	19	"	"	"	136	435	4.0	29.3	14.7	14.4	870	1040	"	"	"

* Large pieces of refractory varying in size between ½" and ¼" were placed on the distributor.

TABLE E III

Test Conditions (Cycle Test 3)

Run No.	Cycle No.	Fuel	Lime-Stone	Limestone Particle Size Range (μ)	Fuel Rate (g/min	Total Air Rate (l/min	Superficial Gas. Vel (ft/sec)	% Stoich. Air	Bed Depth (Static) (inches) Start	Bed Depth (Static) (inches) End	Average Bed Temp. (Absorption) (°C)	Max. Bed Temp. (Regeneration) (°C)	Duration of Absorption (min.)	Distributor	Make-up weight (limestone) (g/cycle)
44/4A	1	US(2.3%S)	BCR(1691	300-3175	197	670	6.1	31.2	15.5	-	860	980	45	Normal*	1200
45/4A	2	"	"	"	231	643	5.7	25.5	-	-	830	1010	"	"	"
46/4A	3	"	"	"	230	653	5.8	26.0	-	-	835	1020	"	"	"
47/4A	4	"	"	"	237	662	6.0	25.6	-	-	850	1030	"	"	"
48/4A	5	"	"	"	241	671	6.1	25.5	-	-	850	1040	"	"	"
49/4A	6	"	"	"	279	644	5.7	21.2	-	-	825	1030	"	"	"
50/4A	7	"	"	"	272	654	5.8	22.1	-	14.2	835	1030	"	"	"
51/4A	8	"	"	"	232	652	5.9	25.8	15.5	-	850	955	"	"	"
52/4A	9	"	"	"	235	621	5.6	24.2	-	-	840	1020	"	"	"
53/4A	10	"	"	"	237	653	5.9	25.3	-	-	850	1030	"	"	"
54/4A	11	"	"	"	248	640	5.7	23.7	-	-	840	1030	"	"	"
55/4A	12	"	"	"	236	637	5.7	24.8	-	-	845	1020	"	"	"
56/4A	13	"	"	"	262	651	5.8	22.8	-	-	835	1040	"	"	"
57/4A	14	"	"	"	224	629	5.7	25.8	-	16.1	845	1030	"	"	"
70/4B	15	"	"	"	178	451	4.0	23.2	13.3	12.4	840	1000	60	Normal*	"
71/4B	16	"	"	"	161	457	4.1	26.0	13.3	13.3	850	1010	"	"	"
72/4B	17	"	"	"	159	452	4.1	26.1	14.0	13.3	850	1040	"	"	"
73/4B	18	"	"	"	161	450	4.1	25.6	14.7	14.7	845	1030	"	"	"
74/4B	19	"	"	"	155	433	3.9	25.6	14.7	14.0	845	1030	"	"	"
75/4B	20	"	"	"	144	467	4.3	29.8	15.3	14.0	870	1040	"	"	"
76/4B	21	"	"	"	145	474	4.3	30.0	15.6	15.5	865	1040	"	"	"
77/4B	22	"	"	"	150	458	4.2	28.0	15.3	14.7	875	1080	"	"	"
78/4B	23	"	"	"	146	460	4.2	28.9	15.3	14.7	860	1040	"	"	"
79/4B	24	"	"	Make-up 600-1400μ	261	658	6.0	23.1	15.3	13.4	855	970	45	"	"
80/4B	25	"	"	"	252	653	5.9	23.8	14.4	14.0	855	1040	"	"	"
81/4B	26	"	"	"	251	690	6.3	25.2	15.3	14.7	865	1040	"	"	"
82/4N	27	"	"	"	252	679	6.2	24.7	15.7	15.3	860	1035	"	"	"
83/4B	28	"	"	"	251	650	5.9	23.8	15.7	15.7	860	1030	"	"	"
84/4B	29	"	"	"	253	662	6.0	24.0	16.7	15.6	860	1040	"	"	"
85/4B	30	"	"	"	251	660	6.0	24.1	17.1	16.5	860	1040	"	"	"
86/4B	31	"	"	"	242	670	6.1	25.4	17.3	16.5	860	1030	"	"	"
58/4A	32	"	"	300-3175	239	645	5.9	24.8	12.7	11.7	860	-	"	"	1600
59/4A	33	"	"	"	247	661	6.0	24.6	14.7	12.4	850	-	"	"	"
60/4A	34	"	"	"	253	675	6.2	24.5	14.7	12.0	865	1030	"	"	"
61/4A	35	"	"	"	265	628	5.6	21.7	14.7	13.1	840	1045	"	"	"
62/4A	36	"	"	"	235	652	6.0	25.1	-	-	865	1030	"	"	"
63/4A	37	"	"	"	257	709	6.5	25.3	-	-	865	-	"	"	"
64/4A	38	"	"	"	257	684	6.3	24.4	-	-	860	1000‡	"	"	"
65/4A	39	"	"	"	245	676	6.2	25.3	16.7	16.0	860	1070‡	"	"	"
66/4A	40	"	"	"	216	640	5.9	27.2	17.3	16.7	870	1020	"	"	"
67/4A	41	"	"	"	237	652	5.9	25.2	18.3	17.4	855	1030	"	"	"

* Large pieces of refractory varying in size between [illegible] and [illegible] were placed on the distributor.

‡ In an attempt to obtain more efficient regeneration, mildly substoichiometric conditions were employed at start of regeneration stage until temperature reached 1000°C. Fuel supply then cut off as [illegible].

TABLE E IV

Test Conditions (Cycle Test 4)

Run No.	Cycle	Fuel	Lime-Stone	Limestone Particle Size Range (μ)	Fuel Rate (g/min)	Total Air Rate (1/min)	Superficial Gas. Vel. (ft./sec.)	% Stoich. Air	Bed Depth (static) (inches) Start	End	Average Bed Temp. (Absorption) (°C)	Max. Bed Temp. (Regeneration) (°C)	Duration of Absorption (min.)	Distri-butor	Make-up weight (limestone) (g/cycle)
68/4A	1	UK(3.03%S)	US(BCR1691)	300-3175	228	635	5.8	25.6	15.5	12.7	855	1010	45	Normal	1200
69/4A	2	"	"	"	227	644	5.9	26.0	12.9	12.9	860	1030	"	"	"
70/4A	3	"	"	"	237	671	6.2	26.0	14.4	12.5	870	1060	"	"	"
71/4A	4	"	"	"	247	677	6.3	25.2	14.0	12.0	875	1040	"	"	"
72/4A	5	"	"	"	248	744	6.9	27.5	13.6	13.3	875	1050	35	"	"
73/4A	6	"	"	"	244	657	6.0	24.7	14.7	13.1	860	1040	"	"	"
74/4A	7	"	"	"	240	678	6.3	25.9	14.7	13.3	880	1040	"	"	"
75/4A	8	"	"	"	252	652	6.0	23.7	15.3	14.0	860	1020	"	"	"
76/4A	9	"	"	"	246	661	6.1	24.7	16.0	14.5	865	1035	"	"	"
77/4A	10	"	"	"	259	663	6.0	23.5	16.5	14.9	850	1040	"	"	"

TABLE E V

Test Conditions (Cycle Test 5)

Run No.	Cycle	Fuel	Lime-Stone	Limestone Particle Size Range (μ)	Fuel Rate (g/min)	Total Air Rate (l/min)	Superficial Gas. Vel. (ft./sec.)	% Stoich. Air	Bed Depth (static) (inches) Start	End	Average Bed Temp. (Absorption) (°C)	Average Bed Temp. (Regeneration) (°C)	Duration of Absorption (min)	Distri-butor	Make-up weight (limestone) (g/cycle)
78/4A	1	US(2.3%S)	US(BCR1691)	300-3175	227	650	5.9	26.3	14.0	13.1	855	1000	45	Normal	2400
79/4A	2	"	"	"	230	690	6.4	27.6	14.7	13.3	870	1030	"	"	"
80/4A	3	"	"	"	222	649	5.9	26.8	15.3	14.0	850	1050	"	"	"
81/4A	4	"	"	"	224	638	5.9	26.1	15.7	15.2	810	1020	"	"	"
82/4A	5	"	"	"	232	697	6.4	27.6	17.3	16.0	860	1025	"	"	"
83/4A	6	"	"	"	247	652	5.8	24.2	18.7	17.3	835	1030	"	"	"
84/4A	7	"	"	"	227	659	5.9	26.6	18.7	18.0	845	1020	"	"	"
85/4A	8	"	"	"	222	665	6.0	27.5	20.0	16.5	850	1030	"	"	"
86/4A	9	"	"	"	214	713	6.6	30.6	17.8	16.8	870	1020	"	"	"
87/4A	10	"	"	"	218	687	6.2	28.9	20.0	18.0	855	1010	"	"	"
88/4A	11	"	"	"	213	648	6.0	27.9	16.0	14.4	870	980	"	"	"
89/4A	12	"	"	"	227	682	6.3	27.6	16.0	14.4	865	1000	"	"	"
90/4A	13	"	"	"	217	649	5.9	27.4	18.0	16.7	855	1010	"	"	"
91/4A	14	"	"	"	222	682	6.2	28.2	18.7	19.3	850	1010	"	"	"
92/4A	15	"	"	"	223	683	6.2	28.1	20.9	20.0	850	1010	"	"	"
93/4A	16	"	"	"	219	682	6.2	28.3	20.0	20.0	860	1020	"	"	"
94/4A	17	"	"	"	222	681	6.2	28.1	20.4	20.0	855	1010	"	"	"
95/4A	18	"	"	"	217	696	6.4	29.4	20.0	20.0	865	1020	"	"	"
96/4A	19	"	"	"	227	682	6.3	27.6	20.0	18.3	875	1020	"	"	"
97/4A	20	"	"	"	234	673	6.3	26.4	18.7	16.7	880	1030	"	"	"

TABLE E VI

Test Conditions (Cycle Test 6)

Run No.	Cycle	Fuel	Lime-Stone	Limestone Particle Size Range (μ)	Fuel Rate (g/min.)	Total Air Rate (l/min.)	Superficial Gas. Vel. (ft./sec.)	% Stoich. Air	Bed Depth (static) (inches) Start	End	Average Bed Temp. (Absorption (°C)	Average Bed Temp.) (Regeneration) (°C)	Duration of Absorption (min.)	Distri-butor	Make-up weight (limestone) (g/cycle)
98/4A	1	US(2.3%S)	US(BCR1691)	300-3175	218	665	6.1	28.0	15.4	14.7	865	980	45	Normal	1200
99/4A	2	"	"	"	231	663	6.1	26.3	16.0	15.4	860	1020	"	"	"
100/4A	3	"	"	"	236	685	6.2	26.6	17.3*	18.3	855	1050	"	"	"
101/4A	4	"	"	"	231	694	6.3	27.6	19.3	18.7	855	1020	"	"	"
102/4A	5	"	"	"	221	656	5.9	27.2	20.5	20.0	850	1020	"	"	"
103/4A	6	"	"	"	219	673	6.2	28.2	20.7	20.0	865	1030	"	"	"
104/4A	7	"	"	"	216	689	6.4	29.3	20.0	20.4	880	1040	"	"	"
105/4A	8	"	"	"	236	679	6.2	26.4	20.0	18.7	800	1040	"	"	"
106/4A	9	"	"	"	231	655	6.0	26.0	20.3	20.0	855	1020	"	"	"
107/4A	10	"	"	"	242	671	6.1	25.4	20.3	19.8	860	1020	"	"	"
89/4B	11	"	"	"	275	661	5.9	22.1	17.0	16.0	835	980	"	"	"
90/4B	12	"	"	"	275	655	5.9	23.4	17.3	16.7	850	1015	"	"	"
91/4B	13	"	"	"	238	666	6.0	25.7	18.7	18.3	850	1020	"	"	"
92/4B	14	"	"	"	238	661	6.0	25.5	18.9	19.5	850	1030	"	"	"
93/4B	15	"	"	"	226	635	5.7	25.8	20.0	19.7	840	1025	"	"	"
94/4B	16	"	"	"	218	642	5.8	27.0	20.0	20.0	850	1025	"	"	"
95/4B	17	"	"	"	217	636	5.8	26.9	20.4	20.4	855	1035	"	"	"

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* 1000 g used bed added to raise height to desired level.

TABLE E VII

Test Conditions (Cycle Test 8)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range (µ)	Fuel Rate (g/min)	Total Air Rate (l/min)	Superficial Gas Vel. (ft/sec)	% Stoich. Air	Bed Depth (Static) (inches) Start	End	Average Bed Temp. (Absorption) (°C)	Max. Bed Temp. Regeneration) (°C)	Duration of Absorption (min)	Distri-butor	Make-up weight (limestone) (g/cycle)
108/4A	1	US(2.3%S)	BCR1691	300-3175	202	659	5.9	29.9	16.0	15.3	835	955	45	Normal	1200
109/4A	2	"	"	"	228	678	6.2	27.3	15.7	15.1	865	990	"	"	"
110/4A	3	"	"	"	230	675	6.2	26.9	16.0	15.3	870	1015	"	"	"
111/4A	4	"	"	"	230	677	6.1	27.0	17.3	16.7	840	1010	"	"	"
112/4A	5	"	"	"	235	686	6.2	26.8	16.7	16.0	855	1030	"	"	"
113/4A	6	"	"	"	241	668	6.0	25.4	16.5	16.0	845	1035	"	"	"
114/4A	7	"	"	"	242	636	5.6	24.2	20.0	19.5	825	1010	"	"	"
115/4A	8	"	"	"	240	664	5.9	25.4	20.7	20.3	830	1000	"	"	"
116/4A	9	"	"	"	242	683	6.2	25.9	21.3	21.3	845	1000	"	"	"
117/4A	10	"	"	"	240	658	5.9	25.2	21.7	21.6	835	1010	"	"	"
118/4A	11	"	"	"	208	650	5.8	28.7	21.3	21.4	840	990	"	"	"
119/4A	12	"	"	"	204	622	5.6	27.8	22.7	22.0	835	1000	"	"	"
120/4A	13	"	"	"	209	676	6.1	29.7	22.7	22.0	850	1010	"	"	"
121/4A	14	"	"	"	197	649	5.8	30.2	22.7	22.9	845	1010	"	"	"
122/4A	15	"	"	"	227	624	5.6	25.2	24.0	22.7	835	1000	"	"	"
123/4A	16	"	"	"	219	641	5.8	26.9	22.7	23.0	840	1010	"	"	"
124/4A	17	"	"	"	232	611	5.4	24.2	24.0	24.7	825	1010	"	"	"
125/4A	18	"	"	"	222	668	6.1	27.6	22.7	23.3	860	1010	"	"	"
126/4A	19	"	"	"	271	655	6.0	22.3	24.0	23.3	860	1030	"	"	"
127/4A	20	"	"	"	266	647	5.8	22.3	24.7	24.0	835	1025	"	"	"
116/4B	21	"	"	"	262	657	5.9	23.0	20.0	18.4	850	1025	"	"	900
117/4B	22	"	"	"	257	632	5.6	22.6	25.3	23.3	825	1010	"	"	"
118/4B	23	"	"	"	256	673	6.0	24.1	24.7	22.4	840	1020	"	"	"
119/4B	24	"	"	"	256	691	6.3	24.8	22.7	22.7	850	1025	"	"	"
120/4B	25	"	"	"	256	668	6.0	23.9	22.9	22.7	840	1030	"	"	"
121/4B	26		"	"	255	693	6.3	24.9	23.3	22.7	855	1010	"	"	"
122/4B	27	"	"	"	256	686	[illegible]	24.6	23.3	22.7	860	1005	"	"	"
123/4B	28	"	"	"	255	690	6.3	24.8	23.3	22.7	865	1040	"	"	"
124/4B	29	"	"	"	256	705	6.5	25.3	22.7	22.7	865	1040	"	"	"
125/4B	30	"	"	"	266	684	6.3	23.6	22.7	22.4	865	1040	"	"	"
126/4B	31	"	"	"	241	679	6.1	25.8	23.3	23.3	850	990	"	"	"
127/4B	32	"	"	"	242	698	6.3	26.5	22.7	22.7	855	1005	"	"	"
128/4B	33	"	"	"	247	660	6.0	24.5	22.7	22.7	850	1030	"	"	"
129/4B	34	"	"	"	246	672	6.1	25.1	22.7	22.4	855	1015	"	"	"

TABLE E VIII

Test Conditions (Cycle Test 8 continued)

Run No.	Cycle No.	Fuel	Lime-Stone	Limestone Particle Size Range (µ)	Fuel Rate (g/min.)	Total Air Rate (l/min)	Superficial Gas Vel. (ft/sec	% Stoich. Air	Bed Depth (Static) (inches) Start	End	Average Bed Temp. (Absorption) (°C)	Max. Bed. Temp Regeneration (°C)	Duration of Absorption (min)	Distributor	Make-up weight (limestone) (g/cycle)
130/4B	35	US(2.[illegible])	US(BCR1691	300-3175	245	665	6.1	24.9	22.7	22.7	860	1030	45	Normal	1200
131/4B	36	"	"	"	255	721	6.6	25.9	22.7	22.0	860	1020	"	"	"
132/4B	37	"	"	"	254	712	6.5	25.5	22.7	22.4	860	-	"	"	"
133/4B	38	"	"	"	257	725	6.6	25.9	22.9	22.6	860	1030	"	"	"
134/4B	39	"	"	"	255	670	6.1	24.1	23.2	22.9	855	1035	"	"	"
135/4B	40	"	"	"	255	666	6.1	24.0	23.2	22.8	855	1040	"	"	"
136/4B	41	"	"	"	227	648	5.8	26.2	22.7	22.7	845	-	"	"	"
137/4B	42	"	"	"	240	709	6.4	27.1	22.7	22.7	855	995	"	"	"
138/4B	43	"	"	"	247	678	6.1	25.2	22.7	22.7	840	1025	"	"	"
139/4B	44	"	"	"	214	658	5.9	28.2	22.7	22.7	830	1010	"	"	"
140/4B	45	"	"	"	202	688	6.2	31.2	23.3	22.7	850	1030	"	"	"
141/4B	46	"	"	"	208	669	6.0	29.5	24.0	23.7	845	1030	"	"	"
142/4B	47	"	"	"	209	696	6.3	30.6	24.8	24.8	850	1030	"	"	"
143/4B	48	"	"	"	209	685	6.2	30.1	25.1	24.0	850	1030	"	"	"
144/4B	49	"	"	"	208	615	5.6	27.1	25.1	24.0	855	1020	"	"	"
145/4B	50	"	"	"	208	658	6.0	27.7	25.3	24.5	860	1040	"	"	"
128/4A	52	"	"	"	218	661	5.9	27.8	23.3	23.3	840	960	"	"	"
129/4A	53	"	"	"	231	678	6.1	26.9	24.4	24.4	835	980	"	"	"
130/4A	54	"	"	"	207	655	5.9	29.0	25.3	25.3	845	1005	"	"	"
131/4A	55	"	"	"	198	679	6.2	31.5	25.3	25.1	860	1010	"	"	"
132/4A	56	"	"	"	219	650	5.9	27.2	25.3	25.3	845	1010	"	"	"
133/4A	57	"	"	"	216	660	5.9	28.0	25.3	25.3	845	1010	"	"	"
134/4A	58	"	"	"	225	681	6.2	28.0	25.3	25.3	850	[illegible]	"	"	"
135/4A	59	"	"	"	222	707	6.4	29.2	25.3	25.3	850	[illegible]	"	"	"
136/4A	60	"	"	"	289	667	6.1	29.3	26.7	25.3	855	1020	"	"	"
137/4A	61	"	"	"	208	687	6.3	30.3	26.0	25.7	860	1020	"	"	"
138/4A	62	"	"	"	214	676	6.8	29.0	26.7	26.0	830	970	50	"	1330
139/4A	63	"	"	"	202	696	6.3	31.6	26.0	26.1	855	1000	"	"	"
140/4A	64	"	"	"	234	678	6.1	26.6	25.3	25.3	840	1020	"	"	"
141/4A	65	"	"	"	218	657	6.0	27.6	-	-	855	1035	"	"	"
142/4A	66	"	"	"	232	693	6.4	27.4	-	-	875	1035	"	"	"
143/4A	67	"	"	"	236	685	6.3	26.6	17.3	17.1	865	1030	30	"	"
144/4A	68	"	"	"	250	683	6.2	25.1	16.4	16.4	855	1030	"	"	800
145/4A	69	"	"	"	284	664	6.0	21.5	16.4	16.0	845	1040	"	"	"
146/4A	70	"	"	"	278	646	5.8	21.3	16.0	16.0	845	1030	"	"	"
147/4A	71	"	"	"	265	678	6.1	23.5	16.3	16.0	850	1030	"	"	"
148/4A	72	"	"	"	259	636	5.9	22.5	14.7	14.0	870	1020	"	"	"
147/4A	73	"	"	"	255	676	6.3	24.3	14.7	14.7	880	1030	"	"	"
148/4A	74	"	"	"	265	649	6.0	22.4	15.3	14.7	865	1060	"	"	"
149/4A	75	"	"	"	267	661	6.1	22.7	16.0	15.7	865	1060	"	"	"
150/4A	76	"	"	"	265	632	6.0	22.6	16.1	16.0	870	1060	"	"	"
151/4A	77	"	"	"	253	650	6.0	23.6	16.7	16.0	875	1060	"	"	"
152/4A	78	"	"	"	274	646	5.9	21.6	17.1	16.3	860	1070	"	"	"
53/4A	79	"	"	"	269	645	5.9	22.0	16.7	16.0	870	1060	"	"	"
.54/4A	80	"	"	"	267	[illegible]	6.0	[illegible]	16.4	16.0	870	1065	"	"	"

TABLE E IX

Test Conditions (Cycle Test 9)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range (µ)	Fuel Rate (g/min	Total Air Rate (l/min	Superficial Gas. Vel. (ft/sec)	% Stoich. Air	Bed Depth (Static) (inches Start	Bed Depth (Static) (inches End	Average Bed Temp(Absorption) (°C)	Max. Bed Temp(Regeneration) (°C)	Duration of Absorption (min	Distri-butor	Make-up weight (limestone) (g/cycle)
159/4B	1	US(2.3%S)	BCR(1691)	300-3175	227	641	5.8	25.9	14.7	13.3	845	1000	30	Normal	800
160/4B	2	"	"	"	223	649	5.9	26.7	13.3	12.7	850	1030	"	"	"
161/4B	3	"	"	"	231	643	5.8	25.5	14.0	14.0	840	1040	"	"	"
162/4B	4	"	"	"	201	650	6.0	29.7	14.7	14.7	870	1010	"	"	"
163/4B	5	"	"	"	214	658	6.0	28.2	15.1	14.7	865	1040	"	"	"
164/4B	6	"	"	"	218	661	6.1	27.8	15.3	15.3	870	1040	"	"	"
165/4B	7	"	"	"	221	658	6.0	27.3	16.0	16.0	860	1040	"	"	"
166/4B	8	"	"	"	219	660	6.0	27.6	16.5	16.4	860	1040	"	"	"
167/4B	9	"	"	"	223	649	5.9	26.7	17.3	16.5	860	1050	"	"	"
168/4B	10	"	"	"	223	646	5.9	26.6	16.3	16.1	860	1040	"	"	"
169/4B	11	"	"	"	223	663	6.1	27.3	16.3	16.0	865	1050	"	"	"
170/4B	12	"	"	"	223	648	6.0	26.7	16.0	15.9	870	1050	"	"	"
171/4B	13	"	"	"	231	653	6.0	25.9	16.0	15.6	860	1035	"	"	"
172/4B	14	"	"	"	229	628	5.7	25.2	15.6	14.7	860	1040	"	"	"
173/4B	15	"	"	"	233	657	6.0	25.9	15.3	15.2	865	1020	"	"	"

TABLE E X

Test Conditions (Cycle Test 10)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range (μ)	Fuel Rate (g/min)	Total Air Rate (1/min	Superficial Gas. Vel. (ft/sec)	% Stoich. Air	Bed Depth (Static) (inches) Start	End	Average Bed Temp. (Absorption) (°C)	Max. Bed Temp. (Regeneration (°C)	Duration of Absorption (min)	Distri-butor	Make-up weight (limestone) (g/cycle)
174/4B	1	US(2.3%S)	BCR(1691	300-3175	118	445	4.0	34.6	20.7	18.4	850*	1000	68	Normal	1000
175/4B	2	"	"	"	101	426	3.9	38.7	20.7	20.1	855*	1020	"	"	"
176/4B	3	"	"	"	110	425	3.8	35.5	22.7	22.0	825*	1020	"	"	"
177/4B	4	"	"	"	103	436	4.0	38.8	22.7	22.7	855*	1025	"	"	"
178/4B	5	"	"	"	111	423	3.8	35.0	22.3	22.0	845*	1030	"	"	"
179/4B	6	"	"	"	124	445	4.0	32.9	22.7	22.7	830*	1030	"	"	"
180/4B	7	"	"	"	117	441	3.9	34.6	22.7	22.7	835*	1040	"	"	"
181/4B	8	"	"	"	117	442	3.8	34.7	22.3	22.0	825*	1000	"	"	"
182/4B	9	"	"	"	117	464	4.1	36.3	22.7	22.3	835*	1030	"	"	"
183/4B	10	"	"	"	131	457	4.1	32.0	22.7	22.3	835*	1030	"	"	"
184/4B	11	"	"	"	136	450	4.0	30.3	22.7	22.9	825*	1050	"	"	"
185/4B	12	"	"	"	118	470	4.3	36.5	23.7	23.3	860*	1030	"	"	"
186/4B	13	"	"	"	119	476	4.3	36.7	22.7	22.7	860*	1040	"	"	"

* Temperature spread between top and bottom of bed 40°C.

TABLE Ξ XI

Test Conditions (Cycle Test 11)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range (μ)	Fuel Rate (g/min)	Total Air Rate (1/min)	Superficial Gas Vel. (ft/sec)	% Stoich. Air	Bed Depth (Static) Start	End	Average Bed Temp. (Absorption)	Max. Bed Temp. (Regeneration)	Duration of Absorption	Distri-butor	Make-up weight (limestone) (g/cycle)
156/4A	1	US(2.3%S)	BCR(1691)	300-3175	150	559	5.2	34.1	18.0	16.8	870	970	55	Normal	-
159/4A	2	"	"	"	170	549	4.9	29.6	19.7	19.3	840	1000	"	"	2000
148/4A	3	"	"	"	175	532	4.8	27.9	20.0	18.7	850	1020	"	"	1000
159/4A	4	"	"	"	185	555	5.0	27.5	20.0	19.7	850	1020	"	"	"
160/4A	5	"	"	"	180	560	5.1	28.5	20.4	20.0	850	1030	"	"	"
161/4A	6	"	"	"	184	526	4.7	26.2	20.0	20.0	846	1020	"	"	"
162/4A	7	"	"	"	187	553	5.0	27.1	20.0	20.0	850	1030	"	"	"
163/4A	8	"	"	"	195	523	4.7	24.6	20.0	20.0	840	1030	"	"	"
164/4A	9	"	"	"	193	544	4.9	25.9	20.0	19.5	850	1030	"	"	"
165/4A	10	"	"	"	166	548	5.0	30.3	19.3	19.3	855	970	"	"	"
166/4A	11	"	"	"	180	565	5.2	28.3	20.0	19.3	865	1010	"	"	"
167/4A	12	"	"	"	194	557	5.0	26.3	20.0	19.6	850	1030	"	"	"
168/4A	13	"	"	"	185	574	5.3	28.5	20.0	20.0	865	1015	"	"	"
169/4A	14	"	"	"	188	536	4.9	26.2	20.0	20.0	855	1020	"	"	"
170/4A	15	"	"	"	189	543	4.9	26.4	20.0	20.0	855	1030	"	"	"
171/4A	16	"	"	"	199	553	5.0	25.5	20.3	20.3	855	1030	"	"	"
172/4A	17	"	"	"	196	549	5.0	25.7	20.0	20.0	855	1030	"	"	"

TABLE E 27

Test Conditions (Cycle Test 12)

Run No.	Cycle No.	Fuel	Lime-stone	Limestone Particle Size Range (µ)	Fuel Rate (g/min	Total Air Rate (l/min	Superficial Gas Vel. (ft/sec)	% Stoich. Air	Bed Depth (Static) (inches) Start	End	Average Bed Temp. (Absorption) (°C)	Max. Bed Temp. (Regeneration) (°C)	Duration of Absorption (min)	Distri-butor	Make-up weight (limestone) (g/cycle)
173/4A	1	UK(3.0%S)	UK	600-1400	125	437	3.9	32.1	16.0	16.2	835	990	68	Normal	750
174/4A	2	"	"	"	115	442	4.0	35.3	16.2	16.0	850	1010	"	"	"
175/4A	3	"	"	"	118	432	3.9	33.6	16.0	16.0	860	1025	"	"	"
176/4A	4	"	"	"	144	438	3.9	27.9	16.0	16.0	840	1040	"	"	"
177/4A	5	"	"	"	143	445	4.0	28.6	15.6	15.3	850	1030	"	"	"
178/4A	6	"	"	"	134	427	3.9	29.2	15.5	15.2	850	1030	"	"	"
179/4A	7	"	"	"	124	449	4.1	33.2	15.1	15.1	860	1000	"	"	"
180/4A	8	"	"	"	133	442	4.0	30.5	15.4	15.1	850	1030	"	"	"
181/4A	9	"	"	"	140	435	3.9	28.5	14.7	14.7	845	1030	"	"	"
182/4A	10	"	"	"	134	418	3.8	26.6	16.0	15.3	860	1010	"	"	"
183/4A	11	"	"	"	143	436	3.9	28.0	16.0	16.0	850	1030	"	"	"
184/4A	12	"	"	"	140	423	3.8	27.7	16.0	16.0	855	1035	"	"	"
185/4A	13	"	"	"	147	[illegible]	3.8	26.5	16.0	15.7	850	1040	"	"	"
186/4A	14	"	"	"	153	438	4.0	[illegible]	16.0	16.0	850	1040	"	"	"

TABLE E XIII

Sulphur Removal Efficiencies (Cycle Tests 1 and 3)

TEST 1						TEST 3						TEST 3 (continued)					
Run No.	Cycle	% Fuel Sulphur Removed *				Run No.	Cycle	% Fuel Sulphur Removed *				Run No.	Cycle	% Fuel Sulphur Removed *			
		15 min	30 min	45 min	60 min			15 min	30 min	45 min	60 min			15 min	30 min	45 min	60 min
52/4B	1	100	97.5	97.8	-	44/4A	1	100	100	100	-	77/4B	22	79.4	79.4	78.8	75.4
53/4B	2	86.6	76.9	76.0	-	45/4A	2	100	100	100	-	78/4B	23	82.8	79.4	75.3	74.8
54/4B	3	70.7	68.9	72.9	-	46/4A	3	85.6	85.6	84.1	-	79/4B	24	58.8	60.2	60.5	-
55/4B	4	78.1	74.7	75.7	-	47/4A	4	-	66.9	67.8	-	80/4B	25	65.2	65.0	59.0	-
56/4B	5	74.8	75.5	72.6	-	48/4A	5	-	80.2	66.4	-	81/4B	26	65.2	61.0	58.2	-
57/4B	6	69.3	70.1	65.9	-	49/4A	6	75.3	73.2	70.4	-	82/4B	27	57.3	61.5	62.0	-
58/4B	7	53.6	59.4	59.6	-	50/4A	7	67.6	68.2	69.0	-	83/4B	28	68.2	65.4	58.8	-
59/4B	8	63.4	71.4	77.5	-	51/4A	8	-	-	-	-	84/4B	29	72.5	-	71.5	-
33/4A	9	-	-	63.1	-	52/4A	9	-	-	-	-	85/4B	30	65.6	68.6	65.6	-
34/4A	10	65.7	67.1	58.8	-	53/4A	10	75.6	61.6	59.6	-	86/4B	31	65.6	68.6	68.2	-
35/4A	11	59.2	64.4	56.5	-	54/4A	11	70.4	65.2	63.9	-	58/4A	32	76.5	69.5	53.8	-
36/4A	12	-	60.0	62.5	-	66/4A	12	69.4	65.0	62.1	-	59/4A	33	51.9	62.1	50.0	-
37/4A	13	62.6	62.4	60.2	-	56/4A	13	59.9	60.7	56.7	-	60/4A	34	70.7	56.1	-	-
38/4A	14	63.6	68/	-	-	57/4A	14	70.0	67.1	67.1	-	61/4A	35	67.9	55.4	54.4	-
60/4B	15	80.0	77.0	70.7	69.2	70/4B	15	80.6	73.6	63.0	59.2	62/4A	36	63.8	55.9	53.1	-
61/4B	16	69.5	69.7	77.6	56.1	71/4B	16	67.2	68.8	66.6	67.6	63/4A	37	75.0	76.6	61.2	-
62/4B	17	76.6	76.3	74.4	71.3	72/4B	17	76.0	74.5	67.1	65.6	64/4A	38	73.8	72.2	65.5	-
63/4B	18	-	76.1	76.1	75.6	73/4B	18	80.0	79.2	76.4	72.8	65/4A	39	73.5	69.0	59.8	-
64/4B	19	68.1	68.1	68.6	68.6	74/4B	19	80.9	76.9	74.7	72.8	66/6A	40	83.4	74.4	61.2	-
-	-	-	-	-	-	75/4B	20	79.0	82.4	78.9	78.2	67/4A	41	77.0	70.7	72.8	-
-	-	-	-	-	-	76/4B	21	82.[illegible]	82.0	72.8	70.3	-	-	-	-	-	-

* Instantaneous Sulphur Removal Efficiency

TABLE E XIV

Sulphur Removal Efficiencies (Cycle Tests 4,5, and 6)

TEST 4						TEST 5						TEST 6					
Run No.	Cycle	% Fuel Sulphur Removed *				Run No.	Cycle	% Fuel Sulphur Removed *				Run No.	Cycle	% Fuel Sulphur Removed *			
		15 min	30 min	45 min	60 min			15 min	30 min	45 min	60 min			15 min	30 min	45 min	60 min
68/4A	1	98.8	98.6	90.2	-	78/4A	1	98.9	98.3	91.2	-	98/4A	1	99.3	98.2	98.0	-
69/4A	2	81.7	75.8	71.0	-	79/4A	2	85.1	73.0	60.8	-	99/4A	2	98.2	98.2	98.1	-
70/4A	3	72.0	58.6	35.4	-	80/4A	3	78.6	75.2	69.5	-	100/4A	3	98.4	95.9	93.2	-
71/4A	4	77.1	54.7	51.5	-	81/4A	4	80.8	73.0	67.0	-	101/4A	4	98.8	96.4	94.1	-
72/4A	5	72.3	61.9	-	-	82/4A	5	85.4	77.8	74.4	-	102/4A	5	98.1	96.4	96.3	-
73/4A	6	80.0	76.7	-	-	83/4A	6	84.2	80.7	77.2	-	103/4A	6	98.1	98.7	96.6	-
74/4A	7	60.0	63.8	-	-	84/4A	7	89.5	84.0	80.6	-	104/4A	7	94.9	89.7	88.9	-
75/4A	8	67.5	57.8	-	-	85/4A	8	86.2	86.2	83.8	-	105/4A	8	96.8	95.2	93.8	-
76/4A	9	-	-	-	-	86/4A	9	-	89.3	79.0	-	106/4A	9	-	94.9	91.0	-
77/4A	10	60.3	59.3	-	-	87/4A	10	90.6	86.8	81.9	-	107/4A	10	96.1	92.5	93.1	-
-	-	-	-	-	-	88/4A	11	-	82.9	66.0	-	89/4B	11	99.0	76.0	80.6	-
-	-	-	-	-	-	89/4A	12	71.9	58.7	61.0	-	90/4B	12	100	96.7	95.8	-
-	-	-	-	-	-	90/4A	13	85.7	85.6	80.2	-	91/4B	13	99.0	90.9	85.3	-
-	-	-	-	-	-	91/4A	14	84.8	85.3	76.9	-	92/4B	14	92.0	89.4	90.0	-
-	-	-	-	-	-	92/4A	15	90.8	93.0	86.7	-	83/4B	15	-	91.5	95.0	-
-	-	-	-	-	-	93/4A	16	-	98.4	95.9	-	94/4B	16	83.4	89.6	85.6	-
-	-	-	-	-	-	94/4A	17	99.5	99.0	99.0	-	95/4B	17	-	-	-	-
-	-	-	-	-	-	95/4A	18	96.7	98.2	-	-	-	-	-	-	-	-
-	-	-	-	-	-	96/4A	19	97.8	92.6	87.6	-	-	-	-	-	-	-
-	-	-	-	-	-	97/4A	20	86.0	76.1	74.9	-	-	-	-	-	-	-

* Instantaneous Sulphur Removal Efficiency

TABLE E-XV

Sulphur Removal Efficiencies (Cycle Test 8)

Cycle	% Fuel Sulphur Removed			* Cycle	% Fuel Sulphur Removed			* Cycle	% Fuel Sulphur Removed			* Cycle	% Fuel Sulphur Removed		
	15 min	30 min	45 min		15 min	30 min	45 min		15 min	30 min	45 min		15 min	30 min	45 min
1	95.4	95.8	99.5	21	99.1	99.1	-	41	86.0	78.0	78.0	61	85.5	94.8	98.8
2	98.2	98.4	99.0	22	100	97.4	97.1	42	73.3	75.6	73.1	62	99.4	96.0	73.4
3	98.4	98.8	98.2	23	99.5	97.0	97.2	43	73.6	74.7	67.1	63	70.6	70.8	68.7
4	98.4	95.2	91.6	24	-	-	-	44	98.5	93.8	87.6	64	61.9	75.5	56.0
5	95.8	-	87.0	25	-	-	-	45	83.4	79.0	79.0	65	79.1	76.3	79.1
6	89.0	82.8	82.8	26	-	-	-	46	80.3	78.0	78.0	66	57.9	47.9	62.4
7	-	97.7	97.7	27	81.3	78.4	76.4	47	78.5	77.1	74.9	67	99.0	99.0	-
8	99.5	98.8	99.0	28	78.7	80.4	78.1	48	72.2	74.6	72.0	68	67.7	64.7	-
9	-	96.0	96.0	29	-	76.9	71.9	49	81.1	79.3	75.7	69	-	-	-
10	100	98.9	98.9	30	85.9	74.7	72.3	50	78.7	77.7	75.1	70	-	-	-
11	100	99.6	98.4	31	-	86.9	78.9	51	85.1	79.7	78.2	71	-	-	-
12	99.5	99.5	99.4	32	84.2	84.7	82.3	52	87.6	80.9	78.1	72	100	78.5	-
13	98.3	99.1	94.8	33	80.7	74.2	70.0	53	80.5	79.9	83.2	73	78.8	78.4	-
14	99.6	99.2	96.7	34	73.7	74.5	74.5	54	75.3	74.0	76.9	74	63.5	69.9	-
16	99.6	93.9	94.2	36	83.7	78.6	74.8	56	71.0	72.8	74.3	76	80.9	77.2	-
17	99.4	99.5	98.7	37	79.0	82.2	83.7	57	75.7	74.5	74.6	77	69.6	59.6	-
18	96.4	89.5	90.2	38	91.0	88.1	85.5	58	72.7	76.6	69.8	78	62.1	59.0	-
19	99.2	99.2	99.2	39	94.7	92.6	91.7	59	74.6	76.2	76.3	79	75.4	73.3	-
20	99.3	97.5	98.0	40	94.0	-	90.1	60	84.5	83.2	80.0	80	77.0	68.0	-

* Instantaneous Sulphur Removal Efficiency

TABLE E XVI

Sulphur Removal Efficiencies (Cycle Tests 9, 10, 11 and 12)

	TEST 9			*	TEST 10			*	TEST 11			*	TEST 12		
	% Fuel Sulphur Removed				% Fuel Sulphur Removed				% Fuel Sulphur Removed				% Fuel Sulphur Removed		
Cycle	15 min	30 min	45 min	Cycle	15 min	30 min	45 min	Cycle	15 min	30 min	45 min	Cycle	15 min	30 min	45 min
1	100	100	-	1	98.9	97.8	100	1	99.2	99.2	99.3	1	98.1	99.6	97.2
2	100	100	-	2	-	99.4	98.5	2	100	99.8	94.0	2	98.4	98.4	98.0
3	-	-	-	3	98.3	98.7	98.6	3	93.5	87.8	82.8	3	100	96.8	-
4	87.2	87.9	-	4	98.6	99.7	99.7	4	88.8	90.5	91.1	4	95.6	93.1	93.1
5	87.2	85.4	-	5	99.5	99.7	94.7	5	-	88.4	88.9	5	97.4	99.3	97.2
6	84.3	79.9	-	6	99.1	94.5	93.4	6	-	-	-	6	98.6	-	95.9
7	85.3	-	-	7	98.2	99.0	98.2	7	-	-	-	7	92.7	94.7	88.7
8	82.1	81.6	-	8	99.0	99.8	99.5	8	-	-	-	8	87.7	91.7	90.6
9	91.6	85.9	-	9	98.4	93.5	84.9	9	-	-	-	9	86.6	89.7	87.7
10	89.1	84.3	-	10	88.6	92.9	86.5	10	89.7	91.4	83.0	10	-	89.1	89.5
11	90.0	86.4	-	11	-	-	-	11	85.8	82.4	75.7	11	87.5	-	86.3
12	89.5	86.8	-	12	94.3	89.7	89.2	12	81.8	79.3	71.7	12	-	87.8	89.7
13	92.4	88.8	-	13	92.9	90.9	88.9	13	88.7	-	81.7	13	91.3	-	89.3
14	-	-	-		-	-	-	14	87.8	81.6	76.7	14	86.0	87.4	86.5
15	-	-	-	-	-	-	-	15	79.3	79.3	76.0	-	-	-	-
-	-	-	-	-	-	-	-	16	82.7	77.0	-	-	-	-	-
-	-	-	-	-	-	-	-	17	81.4	77.3	72.6	-	-	-	-

TABLE E XVII

Cycle Test Regenerations

Test 1

Cycle	Max. SO_2 (% by vol)	Max.Temp. (°C)
1	4.3	1000
2	3.2	965
3	6.0	1020
4	6.7	1030
5	7.5	1040
6	8.9	1030
7	7.5	1050
8	-	1030
9	-	950
10	4.0	1020
11	8.4	1030
12	8.4	1040
13	8.0	1040
14	8.4	1020
15	4.9	1000
16	7.4	1035
17	6.9	1020
18	8.7	1060
19	8.3	1040

Test 3

Cycle	Max. SO_2 (% by vol)	Max. Temp. (°C)
1	5.0	980
2	6.2	1010
3	9.5	1020
4	9.0	1030
5	9.5	1040
6	10.3	1030
7	8.6	1030
8	4.7	955
9	7.2	1020
10	8.2	1030
11	8.6	1000
12	7.4	1020
13	7.4	1040
14	7.2	1030
15	3.1	1000
16	7.2	1010
17	8.4	1040
18	8.6	1030
19	8.4	1030
20	6.8	1040
21	7.4	1040
22	7.4	1080
23	8.0	1040
24	-	-
25	6.9	960
26	9.0	990
27	9.4	1040
28	7.8	1000
29	6.4	1010
30	9.6	1100
31	8.6	1070
32	6.0	1020
33	6.8	1030
34	3.5	970
35	7.3	1040
36	9.0	1040
37	8.6	1035
38	8.6	1030
39	8.0	1040
40	7.8	1040
41	8.2	1030

Test 4

Cycle	Max SO_2 (% by vol)	Max. Temp. (°C)
1	4.3	1010
2	9.4	1030
3	9.0	1060
4	8.6	1040
5	8.2	1050
6	8.6	1040
7	8.0	1040
8	7.8	1020
9	7.9	1035
10	7.2	1040

Test 5

Cycle	Max. SO_2 (% by vol)	Max. Temp. (°C)
1	6.9	1000
2	5.4	[illegible]
3	8.2	1050
4	6.7	1020
5	7.4	1000
6	8.2	1030
7	7.0	1020
8	7.4	1030
9	7.4	1015
10	7.0	1010
11	3.8	980
12	4.3	1000
13	7.2	1010
14	7.4	1010
15	6.2	1010
16	6.6	1020
17	7.0	1010
18	7.2	[illegible]
19	7.4	[illegible]
20	7.4	[illegible]

TABLE E XVIII

Cycle Test Regenerations

Test 6			Test 8									Test 9		
Cycle	Max. SO_2 (% by vol)	Max.Temp. (°C)	Cycle.	Max. SO_2 (% by vol)	Max.Temp. (°C)	Cycle	Max. SO_2 (% by vol)	Max.Temp. (°C)	Cycle.	Max. SO_2 (% by vol)	Max.Temp. (°C)	Cycle.	Max. SO_2 (% by vol)	Max.Temp. (°C)
1	3.1	980	1	4.2	955	28	7.5	1040	55	5.4	1010	1	1.8	1000
2	6.0	1020	2	7.5	980	29	7.1	1040	56	-	1010	2	7.8	1030
3	6.6	1050	3	8.9	1015	30	7.2	1040	57	6.5	1010	3	8.2	1040
4	6.0	1020	4	8.4	1010	31	2.3	990	58	6.3	1020	4	7.8	1010
5	6.6	1020	5	7.9	1030	32	3.9	1015	59	6.0	1010	5	7.2	1040
6	6.6	1030	6	8.7	1035	33	-	1030	60	4.7	1020	6	7.1	1040
7	6.2	1040	7	7.5	1010	34	6.4	1015	61	5.7	1020	7	6.4	1040
8	6.8	1040	8	6.8	1000	35	7.9	1030	62	3.5	970	8	8.2	1040
9	6.6	1020	9	6.8	1000	36	8.5	1020	63	5.0	1000	9	6.2	1050
10	5.0	1020	10	7.0	1010	37	8.3	960	64	7.2	1020	10	5.6	1040
11	3.5	980	11	3.8	1010	38	7.3	1030	65	-	1035	11	6.6	1050
12	6.2	1015	12	8.3	1000	39	7.7	1035	66	7.2	1035	12	6.4	1050
13	7.0	1020	13	6.0	1010	40	7.5	1040	67	5.2	1030	13	6.0	1015
14	6.7	1030	14	6.8	1010	41	-	-	68	6.6	1030	14	6.4	1040
15	6.3	1025	15	7.0	1000	42	6.1	1000	69	7.9	1040	15	6.5	1020
16	5.6	1025	16	7.0	1010	43	6.8	1025	70	6.4	1030			
17	8.6	1035	17	7.1	1010	44	3.9	1010	71	7.2	1030			
			18	6.8	1010	45	5.5	1030	72	2.8	1005			
			19	8.3	1030	46	6.4	1030	73	5.0	1030			
			20	7.9	1025	47	5.9	1030	74	7.5	1060			
			21	6.8	1020	48	6.3	1030	75	6.5	1060			
			22	4.3	1010	49	-	1020	76	7.5	1060			
			23	5.7	1020	50	6.5	1040	77	7.5	1060			
			24	5.7	1025	51	6.8	1030	78	8.3	1070			
			25	6.6	1030	52	2.4	960	79	7.3	1060			
			26	6.8	1000	53	3.9	980	80	8.1	1065			
			27	6.5	1005	54	6.0	1005						

TABLE E XIX

Cycle Test Regenerations

Test 10			Test 11			Test 12		
Cycle	Max. SO_2 (% by vol)	Max. Temp. (°C)	Cycle	Max. SO_2 (% by vol)	Max.Temp. (°C)	Cycle	Max. SO_2 (% by vol)	Max. Temp (°C)
1	3.5	1000	1	1.7	970	1	3.7	990
2	4.2	1020	2	4.5	1000	2	-	1010
3	5.0	1020	3	6.6	1020	3	8.4	1025
4	5.5	1025	4	7.0	1020	4	9.2	1040
5	7.5	1030	5	7.0	1030	5	7.6	1030
6	6.6	1030	6	7.0	1020	6	7.8	1030
7	6.8	1040	7	6.8	1030	7	6.2	1000
8	4.6	1000	8	7.8	1030	8	8.6	1030
9	5.8	1030	9	6.6	1030	9	9.2	1030
10	6.2	1030	10	3.1	970	10	8.6	1010
11	8.7	1050	11	4.8	1010	11	7.7	1030
12	5.8	1030	12	6.6	1010	12	7.8	1035
13	6.6	1040	13	6.6	1015	13	8.4	1040
			14	6.2	1020	14	7.8	1040
			15	5.8	1030			
			16	6.6	1030			
			17	6.6	1030			

TABLE E XX

Metals Pick-up (BCR 1691) (Cycle Test)

Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm
1	860	235	7000
2	1250	260	6900
3	1460	300	6800
4	2040	335	6900
5	1920	295	6900
6	1620	270	6900
7	1600	295	6700
8	1960	300	6800
9	2600	271	7200
10	2400	270	6900
11	2800	265	6900
12	3100	255	6900
13	2300	300	6900
14	3400	290	6900

* After regeneration samples.

Table E XXI

Metals Pick-up (BCR 1691) (Cycle Test 3)

Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm	Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm
1	460	245	7200	22	2580	270	6700
2	920	235	7300	23	2980	270	6500
3	960	270	7100	24	3750	240	6300
4	1030	220	7200	25	3000	260	6400
5	1100	220	7000	24	3600	290	6600
6	1270	205	7200	27	3700	290	6500
7	1920	235	6800	28	3800	310	6700
8	1920	215	6700	29	4200	280	6700
9	1980	210	6700	30	3600	285	6700
10	2030	255	6600	31	3600	280	6500
11	2100	345	6700	32	3900	275	6500
12	2350	215	6600	33	2400	240	6300
13	2660	235	6900	34	3300	275	6800
14	2940	255	7300	35	3500	290	6500
15	2900	300	6700	36	3900	290	6300
16	1990	340	6700	37	4300	315	6500
17	2910	330	6800	38	4900	315	6600
18	2120	295	6700	39	4400	435	6600
19	2720	375	6800	40	4300	370	6600
20	2830	330	6700	41	4100	430	6500
21	2830	300	6700				

* After regeneration samples.

Table E XXII

Metals Pick-up (BCR 1691) (Cycle Test 4)

Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm
1	350	265	6600
2	480	295	7100
3	550	295	6900
4	690	300	6800
5	770	305	7000
6	800	295	6900
7	850	300	6800
8	950	290	6900
9	1000	285	6700
10	840	270	6400

* After regeneration samples.

Table E XXIII

Metals Pick-up (BCR 1691) (Cycle Test 5)

Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm
1	470	250	7000
2	450	280	7000
3	740	285	7000
4	1100	325	6900
5	1040	295	7000
6	1090	290	6800
7	1330	265	6900
8	1400	300	7000
9	1070	325	7000
10	1340	300	6700
11	1280	290	6500
12	1250	290	6700
13	1340	310	6700
14	1740	255	6700
15	1640	290	6800
16	1500	250	6700
17	1380	255	6500
18	1500	250	6700
19	1840	235	6300
20	1650	190	6700

* After regeneration samples

TABLE E XXIV

Metals Pick-up (BCR 1691) (Cycle Test 6)

Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm
1	530	223	6600
2	790	246	6500
3	840	241	6300
4	1060	266	6300
5	1160	238	6500
6	1170	230	6200
7	1220	271	6200
8	1680	259	6200
9	1720	220	6500
10	1940	267	6400
11	2350	263	6900
12	3300	276	6600
13	3000	297	6900
14	2555	262	6900
15	2555	284	7200
16	3200	289	6600
17	3200	283	6400

* After regeneration samples.

TABLE E XXV

Metals Pick-up (BCR 1691) (Cycle Test 8)

Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm	Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm	Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm	Cycle	Vanadium* ppm	Sodium* ppm	Iron * ppm
1	280	550	6500	21	2020	297	5900	41	4900	350	6100	61	3880	254	6100
2	560	318	6800	22	2040	312	5500	42	4700	298	6000	62	3900	326	6100
3	680	320	6800	23	2750	327	5700	43	3900	268	5400	63	3700	286	6300
4	1080	334	7100	24	3120	310	5900	44	3900	296	6200	64	3400	302	6200
5	1000	311	6500	25	3150	313	6300	45	3900	317	5700	65	3900	305	6300
6	1390	302	7200	26	2690	-	-	46	4100	314	6100	66	4300	297	6400
7	1555	308	6600	27	3600	316	6300	47	3900	330	6000	67	3000	320	6000
8	1510	305	6500	28	2250	302	6100	48	4800	313	5900	68	4400	309	6100
9	1780	300	6200	29	2800	303	6300	49	4200	330	6200	69	3400	298	6100
10	2180	295	6200	30	4190	296	6200	50	4100	316	6000	70	4000	288	6000
11	-	-	-	31	2400	273	6100	51	4000	300	5900	71	4300	333	6700
12	-	-	-	32	2900	260	5900	52	4280	270	6000	72	4680	279	6500
13	1630	325	6300	33	3500	287	6300	53	4680	259	6200	73	3900	284	6200
14	-	-	-	34	4200	270	6200	54	3570	277	6100	74	4930	271	6300
15	-	-	-	35	4400	317	6500	55	3500	284	6300	75	3700	274	7000
16	-	-	-	36	2900	266	6100	56	3440	285	6500	76	4690	279	6800
17		-	-	37	3400	279	6100	57	3500	269	6000	77	5100	314	6500
18	-	-	-	38	3800	333	6200	58	4690	283	6000	78	4580	313	6700
19	2440	313	6000	39	3900	301	6100	59	3900	274	6100	79	4160	302	6400
20	-	-	-	40	5700	335	6100	60	3980	284	6400	80	3780	308	6500

* After regeneration samples

Table E XXVI

Metals Pick-up (BCR 1691) (Cycle Test 10)

Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm
1	300	304	6400
2	370	324	6500
3	920	343	7400
4	820	299	6700
5	1070	270	6600
6	1100	268	6300
7	740	268	6000
8	1600	244	6200
9	1500	234	6000
10	1400	234	6200
11	2300	259	6200
12	1900	244	6100
13	2500	238	6300

* After regeneration samples

Table E XXVII

Metals Pick-up (BCR 1691) (Cycle Test 11)

Cycle	Vanadium* ppm	Sodium* ppm	Iron* ppm
1	310	301	6500
2	600	363	6600
3	620	303	6400
4	920	334	6800
5	1160	279	6600
6	1420	304	6400
7	1180	273	6000
8	1470	280	6000
9	2480	292	6200
10	2500	284	6000
11	2060	256	6000
12	2260	287	5900
13	2340	253	5800
14	2140	247	5500
15	2860	255	5700
16	3120	256	5600
17	3060	264	5600

* After regeneration samples

Table E XXVIII

Bed Losses Through Cyclones (Cycle Tests)

Test	Cycles	Losses			Make-up/Lime	
		Total (g)	Rate (g/min)	wt % on fuel	Total (g)	Rate (g/min)
1	1	1103	21.2	9.5	588	11.3
	2-8	1081	3.0	1.5	4536	12.5
	9-14	565	1.9	0.8	3780	12.4
	15-19	1830	4.9	5.0	3780	10.1
3	1-7	2091	5.7	2.7	4536	12.4
	8-14	1361	3.7	1.9	4536	12.3
	15-23	2338	3.7	2.8	6048	9.6
	24-33	1145	2.8	1.3	5292	13.0
	34-41	3472	6.7	3.3	9072	17.5
4	1-10	3119	6.6	3.3	6804	14.4
5	1-10	3075	6.0	3.1	13600	26.5
	11-20	5299	10.3	5.4	10578	20.5
6	1-10	4281	8.3	4.1	8085	15.7
	11-17	2531	7.0	3.3	4533	12.6
8	1-10	5940	11.4	5.6	6800	13.1
	11-20	3348	6.4	3.3	6800	13.0
	21-30	2176	4.2	1.9	5100	9.8
	31-40	2900	5.5	2.7	6234	11.8
	41-43	1539	9.8	5.0	1511	9.6
	44-51	3637	8.7	4.9	5289	12.6
	52-61	3810	7.3	4.0	6800	13.1
	72-83	2324	5.3	2.5	6297	14.4
9	1-15	3727	6.7	3.8	7052	12.8
10	1-7	4381	7.8	4.9	5666	10.1
	8-13	1903	3.9	2.3	4408	9.0
11	1-9	872	1.8	1.8	2647	5.5
	10-17	1536	2.6	1.9	3706	6.3

APPENDIX F

CORRELATION FUNCTIONS FOR SULPHUR REMOVAL EFFICIENCY

In Section II-C-2-h of the text, two equations, 4 and 6, were presented to express sulphur removal efficiency in cyclic batch gasification tests. The basis for these equations is presented here.

As a starting point, a first order reaction in sulphur concentration was used for correlating sulphur removal efficiency in fresh lime bed tests.

$$-\frac{dS}{dt} = k\,S \qquad \text{(F-1)}$$

where S is sulphur concentration remaining in the gas, t is gas residence time/seconds, in bed above fuel injector, and k is a first order rate constant, sec $^{-1}$.

Solution of the first order rate equation gives:

$$d\,\ln S = -\,k\,dt \qquad \text{(F-2)}$$

$$S/So = \frac{1}{e^{(kt)}} = \frac{1}{10^{\frac{(kt)}{2.303}}} \qquad \text{(F-3)}$$

where So is the sulphur concentration that would exist in the gas if no reaction occurred. The equation can be rearranged to express the efficiency of sulphur removal (SRE) where,

$$SRE = 100\left(\frac{So - S}{So}\right) = 100\left(1 - \frac{S}{So}\right) \; ; \qquad \text{(F-7)}$$

$$\text{thus,}\quad SRE = 100\left[1 - \frac{1}{10^{\left(\frac{kt}{2.303}\right)}}\right] \qquad \text{(F-5)}$$

In correlating fresh bed test data, better results were obtained by modifying the equation to use a pseudo residence time θ, rather than the superficial residence time t. The value θ was defined empirically

$$\theta = (t - .051),\ \text{seconds} \qquad \text{(F-6)}$$

A possible reason for the improvement given by using θ rather than t is that a finite time is required for cracking the fuel oil and liberating the sulphur in a form in which it can react with the lime.

With this empirical modification, equation F-5 becomes,

$$SRE = 100 \left[1 - \frac{1}{10^{\left(\frac{k\theta}{2.303}\right)}} \right] \qquad (F\text{-}7)$$

In order to apply equation F-7 to the cyclic test situation where the lime is partly deactivated by prior cycles, the following assumptions were made:

- Rate constant, proportional to lime replacement rate, m.
- Rate constant inversely proportional to quantity of sulphur to which lime is exposed in each cycle, d, the projected sulphur differential.

Thus $$\frac{K}{2.303} = \frac{k}{2.303} \; \frac{m}{m_b} \; \frac{d_b}{d} = k\,a\,\frac{m}{d} \qquad (F\text{-}8)$$

where K is the effective rate constant for the cycled lime, k is the rate constant measured in a batch test with fresh lime, m_b is the equivalent lime replacement rate for a fresh bed test, d_b is the effective potential sulphur differential in a fresh bed test, and

$$a = \frac{d_b}{2.303\, m_b}$$

The first order fresh bed constants were measured at the time at which the bed had reached a sulphur content of 5 wt.%. Thus the effective replacement rate was 100/5 = 20 and the sulphur differential was 5. The constant, a, thus becomes $\frac{5}{2.303 \times 20} = 0.109$,

and equation F-7 becomes,

$$SRE = 100 \left[1 - \frac{1}{10^{\left(.109\, k\, \theta\, \frac{m}{d}\right)}} \right] \qquad (F\text{-}9)$$

Linear regression of the fresh bed test results with U.S. limestone BCR 1691 yielded 33.9 sec $^{-1}$ as the best value for the rate constant, k. Figure 35 of the text compared experimental cycle test data with values given by equation (F-9), text equation (7), and showed that in general the experimental results were slightly more favourable than the equation predicted.

To obtain a better representation of the actual cycle test data, they were regressed with an empirical equation that allowed assignment of different weights to the variables involved. The general form of equation adopted was,

$$SRE = 100 \left[1 - \frac{1}{\text{Exp}\left(\frac{w t^{x} \quad m^{y}}{d^{z}} \right)} \right] \qquad \text{(F-10)}$$

where w, x, y and z are empirical constants.

The linearised form of this equation on which a computerised regression was performed is,

$$\log \left[-\log \left(1 - \frac{SRE}{100}\right) \right] = \log w + x \log t + y \log m - z \log d \qquad \text{(F-11)}$$

where logarithims are to base e.

The cyclic test data matched this equation with a correlation coefficient of 0.94, and yielded the following values for constants.

$w = 9.285$
$x = 0.6$
$y = 0.95$
$z = 1.9$

The data therefore show that sulphur differential with an exponent of 1.9 has a much larger effect on sulphur removal than gas residence time with an exponent of 0.6

The final empirical equation is;

$$SRE = 100 \left[1 - \frac{1}{\text{Exp}\left(\frac{9.258\, t^{.6} \quad m^{.95}}{d^{\,1.9}} \right)} \right] \qquad \text{(F-12)}$$

Figure F-1 shows the match between experimental and computed efficiencies using equation F-12 which corresponds with text equation 6.

Values predicted using equation F-12 are listed in Tables F-I to III.

Table F I

Prediction of Lined Out Efficiencies from Cycle Test Results

Residence Time (sec)	Make-Up Rate (Wt.CaO/wt.S)	Potential Sulphur Differential (wt.%)	Lined Out Efficiency (%)
0.1	1.8	2.0	66.707
0.2	1.8	"	81.026
0.3	1.8	"	87.951
0.4	1.8	"	91.888
0.5	1.8	"	94.324
0.6	1.8	"	95.916
0.1	0.8	"	39.930
0.2	0.8	"	53.708
0.3	0.8	"	62.492
0.4	0.8	"	68.775
0.5	0.8	"	73.537
0.6	0.8	"	77.280
0.1	1.3	"	55.414
0.2	1.3	"	70.497
0.3	1.3	"	78.864
0.4	1.3	"	84.193
0.5	1.3	"	87.839
0.6	1.3	"	90.045
0.1	2.3	"	75.089
0.2	2.3	"	87.719
0.3	2.3	"	93.076
0.4	2.3	"	95.797
0.5	2.3	"	97.321
0.6	2.3	"	98.232

Table F II

Prediction of Lined Out Efficiencies from Cycle Test Results

Residence Time (sec)	Make-Up Rate (Wt.CaO/wt.S)	Potential Sulphur Differential (wt.%)	Lined Out Efficiency %
0.1	1.8	1.5	85.065
0.2	1.8	1.5	94.350
0.3	1.8	1.5	97.423
0.4	1.8	1.5	98.700
0.5	1.8	1.5	99.[illegible]99
0.6	1.8	1.5	99.603
0.1	0.8	1.5	58.568
0.2	0.8	1.5	73.594
0.3	0.8	1.5	81.647
0.4	0.8	1.5	86.632
0.5	0.80	1.5	89.958
0.6	0.80	1.5	92.286
0.1	1.3	1.5	75.254
0.2	1.3	1.5	87.881
0.3	1.3	1.5	93.191
0.4	1.3	1.5	95.880
0.5	1.3	1.5	97.382
0.6	1.3	1.5	98.276
0.1	2.3	1.5	90.921
0.2	2.3	1.5	97.337
0.3	2.3	1.5	99.011
0.4	2.3	1.5	99.583
0.5	2.3	o.5	99.809
0.6	2.3	1.5	99.907

Table F III

Prediction of Lined Out Efficiencies from Cycle Test Results

Residence Time (sec)	Make-Up Rate (Wt.CaO/wt.S)	Potential Sulphur Differential (wt.%)	Lined Out Efficiency %
0.1	1.8	3.0	39.855
0.2	1.8	3.0	53.621
0.3	1.8	3.0	62.403
0.4	1.8	3.0	68.686
0.5	1.8	3.0	73.451
0.6	1.8	3.0	77.198
0.1	0.8	3.0	20.990
0.2	0.8	3.0	29.955
0.3	0.8	3.0	36.448
0.4	0.8	3.0	41.611
0.5	0.8	3.0	45.911
0.6	0.8	3.0	49.593
0.1	1.3	3.0	31.160
0.2	1.3	3.0	43.123
0.3	1.3	3.0	51.249
0.4	1.3	3.0	57.376
0.5	1.3	3.0	62.242
0.6	1.3	3.0	66.234
0.1	2.3	3.0	47.350
0.2	2.3	3.0	62.070
0.3	2.3	3.0	70.896
0.4	2.3	3.0	76.893
0.5	2.3	3.0	81.238
0.6	2.3	2.0	84.515

CORRELATION BETWEEN EXPERIMENTAL AND CALCULATED LINED OUT EFFICIENCIES (USING EQUATION F12)

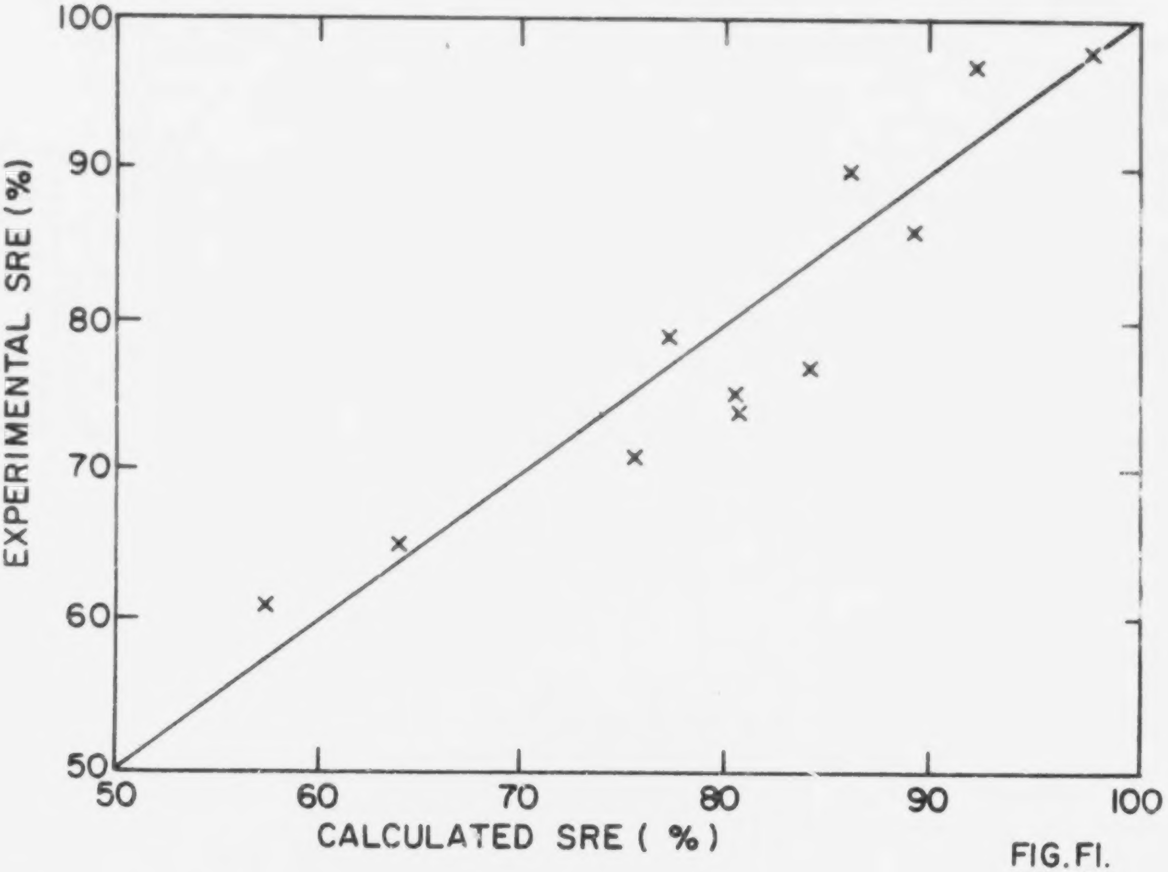


FIG. F1.

APPENDIX G

Design Basis for the Pilot Plant Gasifier System

A. The Gasifier

The size of the experimental gasifier was determined by the decision to adopt dimensions for the regenerator virtually identical with those of the batch unit. This decision was taken on the grounds that this was a proven design which had operated under regeneration conditions in a satisfactory manner at the gas velocities which were envisaged.

Since experience had also shown that the SO_2 concentration in the regenerator tail gas would be in the region of 10% by volume, the sulphur throughput of the unit, under conditions which had actually been used, was limited to 8.6 lbs/hr. In the case of a 2.2% by wt. sulphur fuel the fuel throughput was therefore restricted to 391 lbs/hr. giving an output of 7.1 MM Btu. gross or 6.68 MM Btu. net, per hour. At 20% of stoichiometric air and a gas velocity of 4 ft/sec. at 900°C this indicated a cross-sectional area for the gasifier of 4.8 sq. ft.

At an early stage in the design of the installation it was decided to use the gasifier to fire a conventional packaged hot water boiler. It was considered advantageous to show that the gasifier could be used in conjunction with commercially available equipment, and the installation lent itself readily to the employment of flue gas recycle as a means for controlling the temperature of the gasifier bed. The boiler also provided a suitable test chamber for the development of the gas burner. This was a necessary requirement since no burner handling this type of gas is in current commercial use.

A schematic layout of the completed plant is shown in text Figure 9. The pressurised hot water boiler is provided with a heat exchanger which transfers the heat generated to an atmospheric pressure evaporative cooler. The heat transfer system is completely automatic and maintains a constant boiler water temperature regardless of variations in the heat input. As can be seen it was judged advantageous to set the gasifier in a 6 ft deep pit in order to simplify the ducting between it and the burner.

The actual geometric configuration of the gasifier was very largely influenced by the decision to build it as a monolithic structure in refractory concrete. A major advantage of this method of construction is that it allows heat to be conserved in the transfer system between the two fluidised beds. It also allows the beds to be positioned very close to each other, so reducing the overall size of the unit. Disadvantages arise from the fact that the unit is very difficult to modify once the concrete has been poured. In addition considerable reliance is placed on the durability of the concrete, since the development of cracks in the walls between the gasifying and regenerating beds could adversely affect performance.

This precise form of construction is not recommended for general use but was considered appropriate to the case of a small experimental unit of limited life. Every effort was made to obtain the best advice available in choosing the refractory material and in developing a detailed design for the refractory block.

The geometrical configuration of the gasifier was determined largely by the need to accommodate cyclones for the desulphurised gas stream. It was found that the choice of a single cyclone resulted in an awkward layout and increased the overall height, and therefore cost, of the unit. For this reason two cyclones were used in parallel to give the layout shown in text Fig.12. Since the cyclones consisted merely of holes cast in the refractory block it cost very little more to provide two small cyclones in place of one large one, and the extra expenditure on cores was more than offset by the saving due to the reduction in the overall size of the gasifier.

The cyclones were positioned one on each side of the regenerator. This allowed the use of a rectangular gasifier bed and resulted in a compact design. Another advantage of this layout was that it resulted in a simple ducting system for the transfer of bed material. Once the cyclones had been positioned either side of the regenerator it seemed reasonable to make use of the bed transfer system in order to drain them of fines. In the arrangement which was adopted half of the fines from the gasifier are returned to the gasifier bed, the remainder being fed to the regenerator. The regenerator cyclone, which is an external fitting, is used to drain fines from the unit.

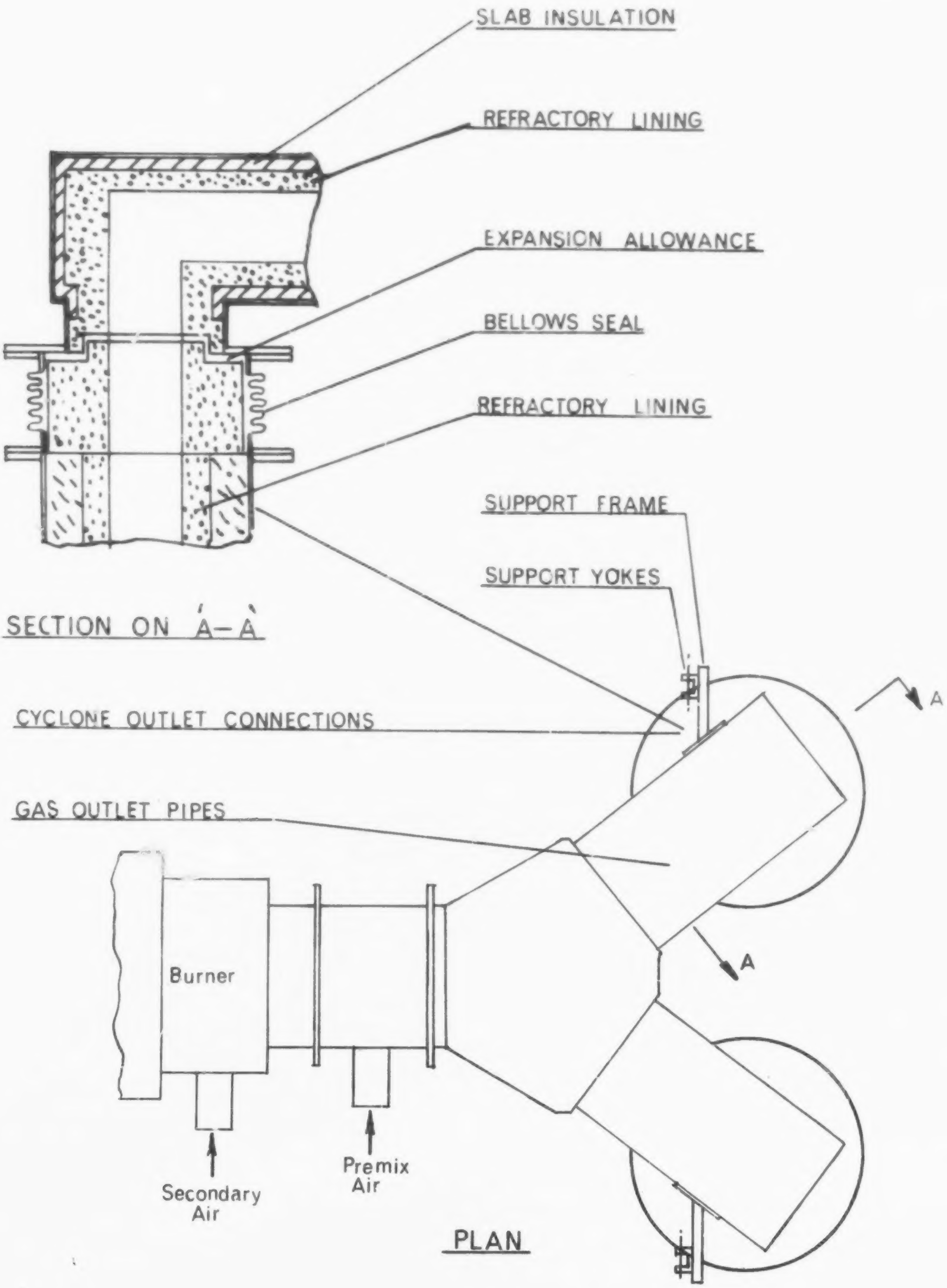
Another question which had to be considered was the manner in which the gasifier should be brought to its working temperature. The small batch units are initially fired with LPG beneath their distributors. This procedure results in an excessive pressure drop across the distributor under starting conditions, and also imposes thermal stresses on its structure. For these reasons it was decided to position the starting burner above the distributor of the gasifier in such a way that it acted as a submerged combustor in the fluidised bed. Since with this arrangement the underside of the distributor was always cold the distributor could be constructed of steel.

In order to facilitate any necessary repairs or alterations to the distributor it was designed for easy removal , and adequate ground clearance was allowed beneath the gasifier for this purpose.

Since refractories are not very good thermal insulators it was necessary to incorporate a shell of insulating material between the monolithic gasifier and its external mild steel casing. The large temperature difference under working conditions between the refractory block and its casing, resulting from this arrangement, introduced a differential thermal expansion problem. In order to allow for the "growth" of the refractory block in the vertical direction it was found necessary to "float" the lid of the gasifier on the refractory block instead of bolting it to the flange of the casing. Differential expansion in the horizontal sense was dealt with by leaving a gap between the refractory block and its insulating shield.

The need to accommodate differential thermal expansion dictated the designs of the offtakes from both the gasifier and the regenerator. The simplest possibility which presented itself was to run these off-takes through the lid of the gasifier so that they could "float" with the refractory block and not be constrained in any way by the steel casing.

The design of the ducting from the gasifier cyclones to the burner is shown in Fig. G-1. The main component is a Y shaped bifurcated duct which runs horizontally and incorporates a 90° elbow for each cyclone outlet connection. This duct is supported at three points. It is suspended close to each cyclone connection by a hanger attached to a frame fixed to the gasifier casing, whilst the common duct rests on a roller which allows horizontal movement along the axis of the boiler fire tube. The connections between the bifurcated duct and the cyclone outlets are spigoted and incorporate sufficient clearance to allow for the vertical expansion of the gasifier block. Gas-tightness is ensured by the use of thin walled steel bellows which enclose the spigoted joints.


SLAB INSULATION
REFRACTORY LINING
EXPANSION ALLOWANCE
BELLOWS SEAL
REFRACTORY LINING
SUPPORT FRAME
SUPPORT YOKES
SECTION ON A–A
A
CYCLONE OUTLET CONNECTIONS
GAS OUTLET PIPES
A
Burner
Secondary Air
Premix Air
PLAN

CYCLONES TO BURNER MANIFOLD

Fig. G-1

B. Design Flow Sheet

Figure G-2 is a schematic flow diagram of the gasifier-regenerator system showing the mass flow rates of the major streams. The basic assumptions on which this material balance is based are listed below.

Sulphur throughput	8.6 lb/hr
Fuel oil sulphur content	2.2 wt. %
Air/fuel ratio	20% of stoichiometric
US limestone	1691
% sulphur in lime* from gasifier	5%
% sulphur in lime* from regenerator	1%
Lime make-up rate	1 mole CaO/mole S fed
SO_2 in regenerator off gas	10 mol. %
Carbon on gasifier lime*	0.3 wt. %

* Lime taken as Ca + O + S in stone

MATERIAL BALANCE AND FLOWS
CONTINUOUS CAFB PILOT PLANT
(ALL FLOWS IN POUNDS/HR)

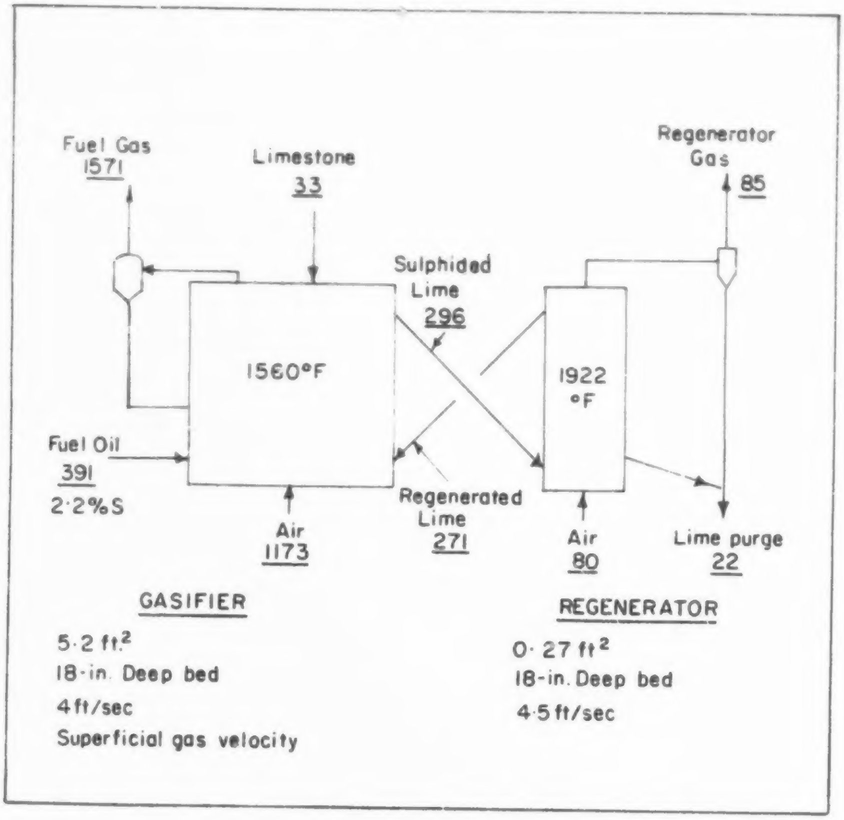

Fuel Gas
1571
Limestone
33
Regenerator Gas
85
Sulphided Lime
296
1560°F
1922 °F
Fuel Oil
391
2·2%S
Air
1173
Regenerated Lime
271
Air
80
Lime purge
22
GASIFIER
5·2 ft.²
18-in. Deep bed
4 ft/sec
Superficial gas velocity
REGENERATOR
0·27 ft²
18-in. Deep bed
4·5 ft/sec

Fig G-2

C. Principles of Gasifier Control

The results of the batch tests indicated that the gasifier would function best at a temperature in the region of 850°C. It was also found that the temperature of the regenerator bed should lie in the region of 1050°C. Since the regeneration reaction is exothermic and since the air and stone entering the regenerator are heated to the operating temperature by the heat released, it is evident that temperature control may be obtained by varying the rate at which stone circulates between the gasifier and the regenerator. A consideration of the chemical reactions involved indicates that the regenerator temperature is, to some extent, self regulating so that the system is inherently stable.

Calcium sulphide is oxidised by two reactions which may be regarded as successive stages of oxidation. The combination of $1\frac{1}{2}$ moles of O_2 with a mol of CaS results in the formation of $CaSO_3$ which is either oxidised by an additional $\frac{1}{2}$ mol of O_2 to $CaSO_4$ or else decomposes to yield a mol of CaO and a mole of SO_2. Complete oxidation to $CaSO_4$ yields 220 Kcal/g mol of CaS whilst decomposition yields only 109.5 Kcal/g mol.

There is a thermodynamic limit to the concentration of SO_2 which can exist in equilibrium with lime at any given temperature. Any excess SO_2 will recombine to form $CaSO_3$ which in the presence of O_2 will be oxidised to $CaSO_4$. Since this temperature/ equilibrium concentration relationship is known,(a) it is possible to compute the heat released per mole of CaS at any given temperature under conditions in which there is no oxygen in the gas leaving the process. This has been done in Table G-I.

Table G-I.

Heat released per mole of CaS When oxidised in the absence of excess oxygen

Regenerator Temp. °C	Equilibrium Press. SO_2 Ats	Mol. Ratio SO_4/SO_3	Mols of O_2/Mol CaS	Heat Released Kcal/Mol CaS
970	0.053	1.62	1.81	177
980	0.064	1.20	1.77	169
994	0.082	0 74	1.71	156
1000	0.091	0.58	1.68	150
1004	0.097	0.48	1.66	145

In addition to CaS the stone leaving a CAFB gasifier also carries about 0.3% by weight of carbon. This carbon will largely be oxidised to CO_2 under continuous running conditions but its effect on the heat released will depend on the carbon/sulphur ratio. In a typical situation the stone might contain 3.0%S and 0.3%C by weight. This is equivalent to 0.267 moles C/mol CaS. On this basis the last two columns of Table I may be adjusted to allow for the additional oxygen consumption and heat release, 93.5 Kcal/mol.C.

(a) See Appendix A for thermodynamic values.

O_2 consumption and heat released per mole of CaS, allowing for carbon at 0.267 moles/mol.

Regenerator Temp. °C	Molar Ratio O_2/CaS	Heat Released Kcal/Mol.CaS
970	2.04	195
980	2.002	186
994	1.932	170
1000	1.897	163
1004	1.872	157

In a large unit only a very small fraction of the heat released will be lost through the wall of the regenerator and, over the temperature range considered, this loss will be almost constant. As a first approximation it may be assumed that all of the heat produced goes to raise the temperature of the stone and the air. If the heat capacities of the stone and the air are both equal to 0.24 KCal/Kg.°C, and if the gasifier temperature is 850°C, the regenerator temperature 1000°C and the air temperature 15°C then the sulphur content of the stone for temperature equilibrium may be computed as follows.

heat released by oxidation/mole CaS = heat absorbed by air + heat absorbed by stone

$$\left[\text{kg air}\right] \times \left[C_p\right] \times \left[\Delta T\right]$$

$$163 \text{ k cal} = \frac{1.897 \times 29}{1000 \times .21} \times .24 \times 985$$

$$+ \left(\frac{\text{kg stone}}{\text{mol S}}\right) \times .24 \times 150$$

$$163 \text{ kcal} = 62 + 36 \times (\text{Kg stone/Mol S})$$

$$\text{Kg Stone/Mol S} = 2.8$$

$$\text{wt \% S in Stone} = \% \ \frac{32 \times 100}{2.76 \times 100} = 1.14\%$$

Of this oxidised sulphur, the amount released as SO_2 is 0.845% by weight on stone. If the gasifier temperature is held constant and the regenerator temperature is allowed to vary the results in Table G III are obtained on the assumption that all the oxygen in the regenerator air supply is utilised.

Table G-III

Oxidised Sulphur Content of Stone
and Sulphur loss per pass

Regenerator Temp. °C	Oxidised Sulphur % wt. Stone	Sulphur lost as SO_2 % wt. stone
970	0.71	0.32
980	0.82	0.43
994	1.03	0.70
1000	1.14	0.84
1004	1.26	0.99

If excess air is used the tendency will be to form more sulphate and less SO_2, more heat will be generated whilst sulphur will be conserved. The effect of this will be to lower the concentration of SO_2 observed at any given temperature and to transfer oxygen from the regenerator to the gasifier.

The inherent stability of the system is indicated by Table G-III which shows that as the regenerator temperature rises the amount of sulphur lost from the system as SO_2, for a given stone throughput, increases very rapidly. If the sulphur input is constant this will result in a rapid approach to the temperature at which the SO_2 output matches the sulphur input. Variation of the stone transfer rate will allow this equilibrium temperature to be varied. It is therefore desirable to have an accurate and reliable means for controlling bed transfer rate in order to ensure that the equilibrium temperature is low enough to avoid dead burning of the stone and the imposition of undue thermal stresses on the structure of the gasifier.

D. The Design of the Bed Transfer System

The system which is used to effect the transfer of bed material between the two reactors employs gravitational dense phase transfer. Each bed container is provided with a catchment pocket slightly above the surface of the bed, and each of these pockets communicates via an almost vertical duct with the side wall of its neighbour at a point slightly above the level of the distributor. There is a short horizontal section at the bottom of each duct so that in the absence of any external stimulus the two transfer ducts simply fill with static bed material. An arrangement of this type is shown in Figure G-3 which gives a schematic layout of the first bed transfer test rig.

TEST RIG FOR BED TRANSFER TRIALS

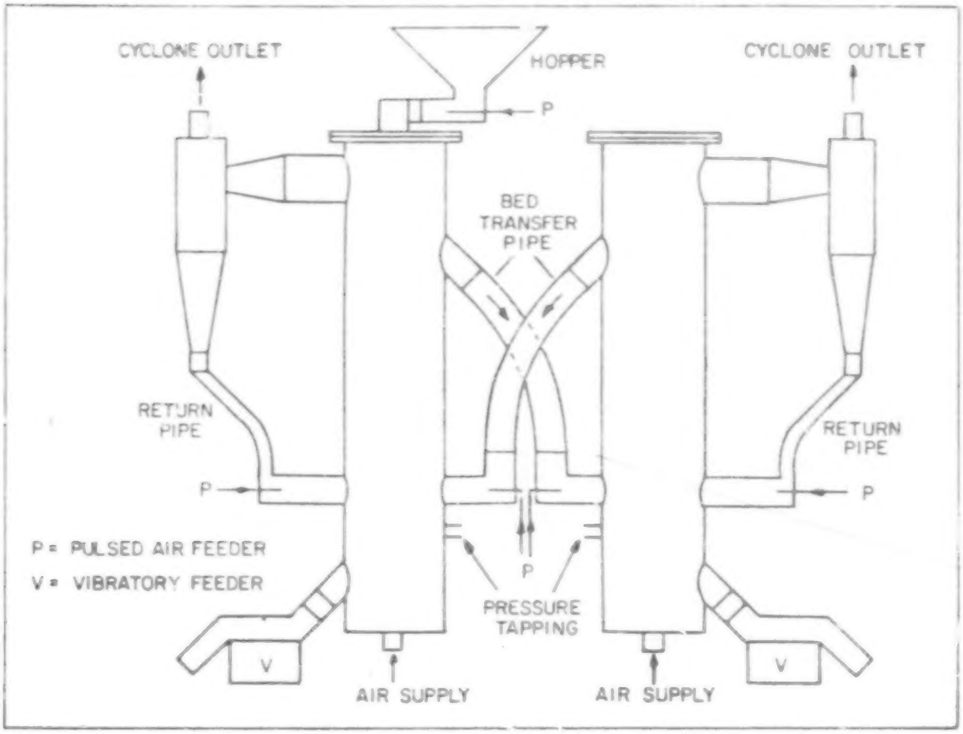

CYCLONE OUTLET
HOPPER
P
CYCLONE OUTLET
BED
TRANSFER
PIPE
RETURN
PIPE
RETURN
PIPE
P
P
P = PULSED AIR FEEDER
V = VIBRATORY FEEDER
P
PRESSURE
TAPPING
V
V
AIR SUPPLY
AIR SUPPLY

Fig.G-3

Each horizontal connection is provided with a suitable tube running parallel with its axis. When gas is introduced via one of these tubes the particles in the horizontal duct are fluidised and expelled, being replaced by gravitational flow down the vertical duct. As soon as the gas ceases to flow so do the solids. By pulsing that gas flow and varying the frequency of the pulses it is possible to control closely the rate at which bed material is transferred. Very little gas is required to operate the system. In the original test rig it was possible to shift about two pounds of bed material per cubic foot of gas. In view of the possible variation of sulphur content throughout the depth of the bed, the transfer of material from the top of each bed to the bottom of the other is advantageous from an operational point of view. The dense packing in each duct, together with the continuous gas bleed which is required to prevent blockage of the activating tubes, ensures that the leakage of gas between compartments is minimal.

The layout of the bed transfer ducts was largely dictated by the geometry of the purifying and regenerating compartments. It was not possible, however, to produce a detailed design on theoretical grounds. There was no way of predicting the dimensions and positions of the catchment pockets which would provide satisfactory results for example, and it would have required an excessive degree of confidence in the outcome to have simply guessed the dimensions of the transfer ducts. It was, therefore, decided at an early stage that the best procedure would be to build a full scale cold model so that these finer points could be settled experimentally. A view of the cold model is shown in Figure 5-4.

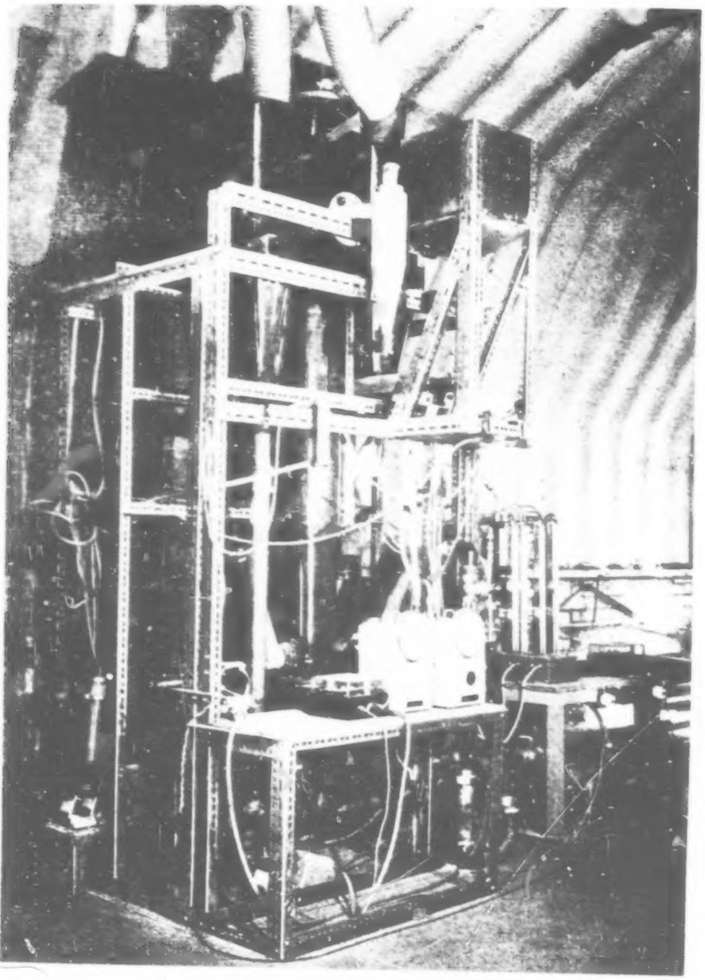


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Fig. 5-4

The greater part of the structure was constructed of thin steel sheet, but the bed transfer system was made of transparent plastic so that the behaviour of the flowing solids could be observed. Plastic construction also lent itself to rapid modification.

The bed material which was used to test the cold model was a crushed brick having the same size distribution and bulk density as the lime which was normally used in the batch units.

Trouble was in fact encountered with the original design in which the fines from the main cyclones were led directly into the bed transfer ducts through branch pipes. It was found that the prevailing flow of gas up the transfer ducts from the gas injectors carried fines into the interstices between the coarser material up stream of the cyclone branches and that this resulted in bridging. This difficulty was overcome by adopting the configuration shown in Figure G-5.

SUCCESSFUL CYCLONE FINES RETURN ZONE

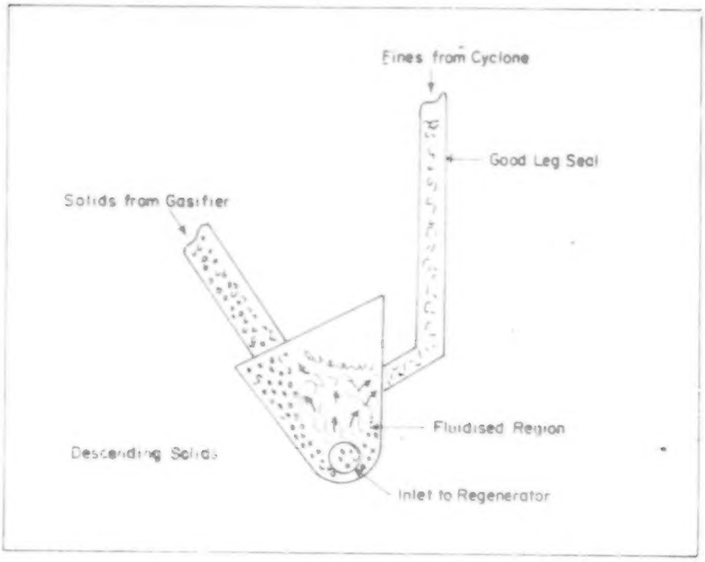


Fig. G-5

The transfer duct was enlarged at the point at which it changed direction horizontally to form a chamber having a vertical face which was swept by gas bubbles rising from the injector. The shape of this chamber was such that a free surface of bed material existed within it, so that the gas bubbling up the verticle face could only continue on up the duct by re-entering this surface which therefore acted as a trap for fines.

The fines from the cyclone entered the entrainment chamber through its bubble swept face, and the mouth of the duct from which they drained was positioned so that it was partially open to the voidage above the free bed surface. This system was found to function well, and it was possible to observe entrained pockets of fines being carried towards the delivery ports by the coarser bed material.

Tests made on the cold rig showed that the system could circulate bed material at a maximum rate of 700 lbs/hr. A cold check on the completed gasifier confirmed this result. Indications are that the transfer system consumes much less nitrogen under hot conditions than it did cold when the requirement was about 1 cubic ft for every two pounds of bed material transferred.

In the control system which has been adopted the pulse rate of the gasifier to regenerator transfer line is controlled by a simple variable proportional band two term controller, according to the deviation of the regenerator temperature from a chosen set point. The operation of the return line from the regenerator to the gasifier is controlled by a pressure switch actuated by the pressure drop across the regenerator bed.

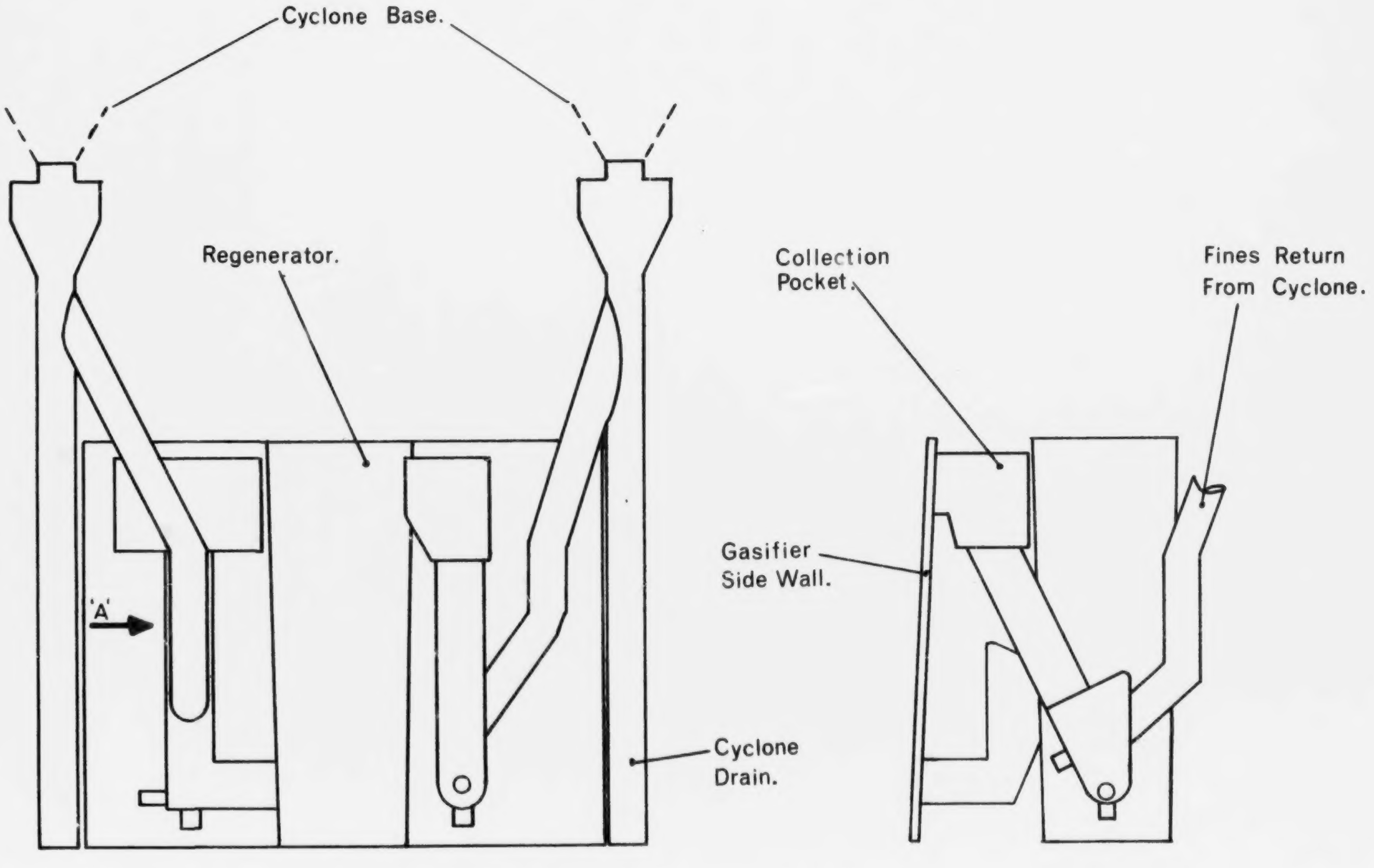
It has been found that this system is somewhat sensitive to differences in pressure between the gas spaces of the regenerator and gasifier and that it is necessary to ensure that these pressures are within a few inches water gauge (w.g.) of each other. A butterfly valve was fitted to the regenerator offtake for this purpose and was adjusted manually. An automatic control loop will be installed for this purpose in future runs.

E. Gasifier Construction

Construction of the gasifier unit itself consisted of bolting together the base and prelined steel sides, building up the cores and molds which form the interior voids, pouring the refractory concrete interior and assembling the lids and distributors.

The base is mounted on concrete piers in a pit which houses the gasifier unit. Each of the eight side plates was lined with a 3-inch thick layer of 50 lb/cu.ft castable insulation which was poured in place on the ground before erection. All penetrations through the walls had been precast in this insulating concrete. The bottom four side sections were assembled on the base and lined with a $\frac{1}{4}$-inch thick blanket of expanded polystyrene sheet to provide the expansion gap. The wooden lower sections of gasifier and regenerator cores were set in place along the plexiglas cores for the transfer lines and catch pockets. The plastic sections were filled with paraffin wax to provide rigidity.

Figure G-6 is a drawing of the plastic patterns. Two views of the plastic form-work installed are shown in Figure G-7 and in text Figure 13.


Cyclone Base.
Regenerator.
'A'
Cyclone Drain.
Collection Pocket.
Fines Return From Cyclone.
Gasifier Side Wall.

VIEW ON ARROW. 'A'.

PERSPEX PATTERN FOR TRANSITION DUCTS

Fig.G-6

PHOTOGRAPH OF GASIFIER UNDER CONSTRUCTION

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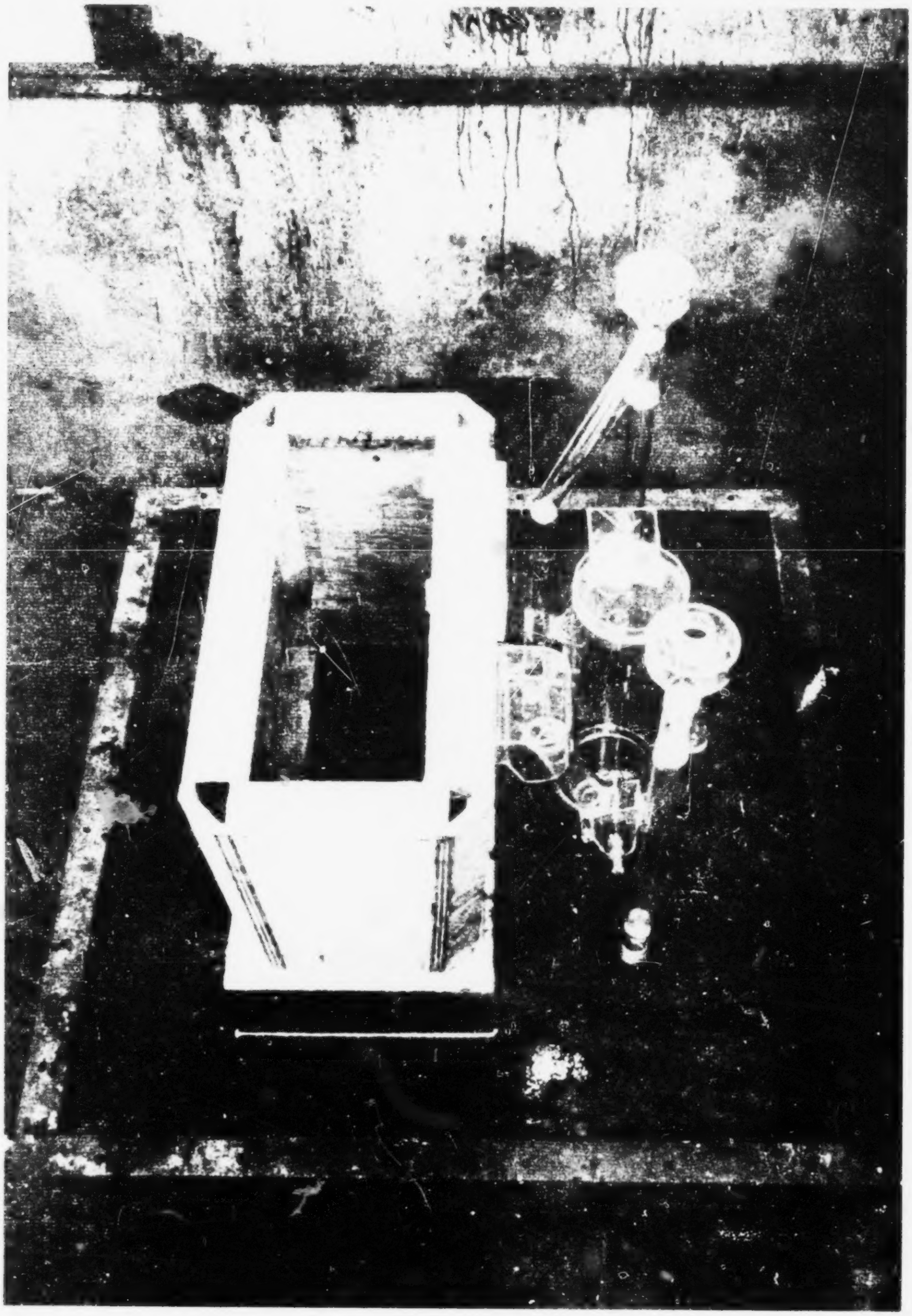

Fig. G-7

The metal inserts for fuel injection, thermocouples, pressure taps, sampling etc. were installed and grease coated to provide for easy removal.

The first three lifts of refractory concrete were then poured. The material was Durax C.1600 Aluminous Hydraulic Concrete obtained from G.R. Stein Refractories Ltd., poured and vibrated to maximum density.

Table G-IV lists properties of this concrete.

Table G-IV

Properties of Durax C.1600 Concrete

Maximum Service Temp.	1600°C
Maximum Grain size	3/8 in.
Al_2O_3 content	49.9 %
Fe_2O_3 content	0.6 %
Density, Cast and Dried	142 lb/ft^3
Density after firing to 1600°C	136 lb/ft^3
Compression strength after firing	6727 lb/in^2
Modulus of Rupture after firing	808 lb/in^2
Thermal Conductivity at 600°C	7.615 $Btu/hr.ft^2.°F/in.$

When the first lifts had set, the upper four side plates were bolted into place, polystyrene blankets, upper cores and inserts placed, and the final three lifts of concrete poured. Precast concrete forms were used in the start-up burner entrance, for the regenerator outlet section, and for the cyclone outlets. Stainless steel inserts formed the cyclone outlet tubes. Final assembly steps consisted of pouring a layer of insulating cement on the top, seating the precast lid, installation of refractory fibre packing in the joints, and bolting on the cover plate. Figure G-8 shows a photograph of the unit after the first three lifts had been poured. A photograph of the completed installation is shown in Figure G-9.

PHOTOGRAPH OF GASIFIER WITH LOWER CONCRETE LIFTS POURED

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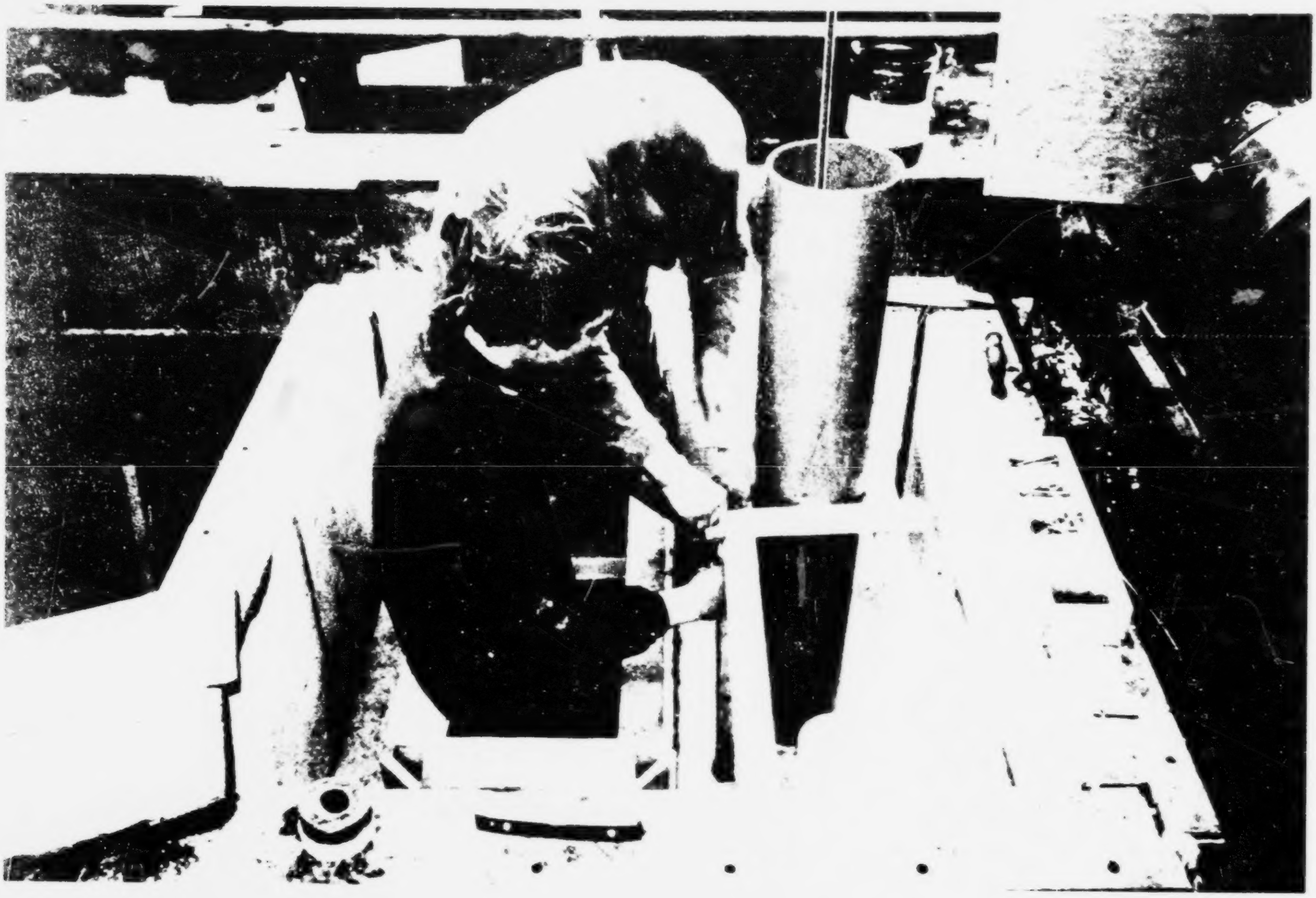

Fig G-8

PHOTOGRAPH OF COMPLETED GASIFIER

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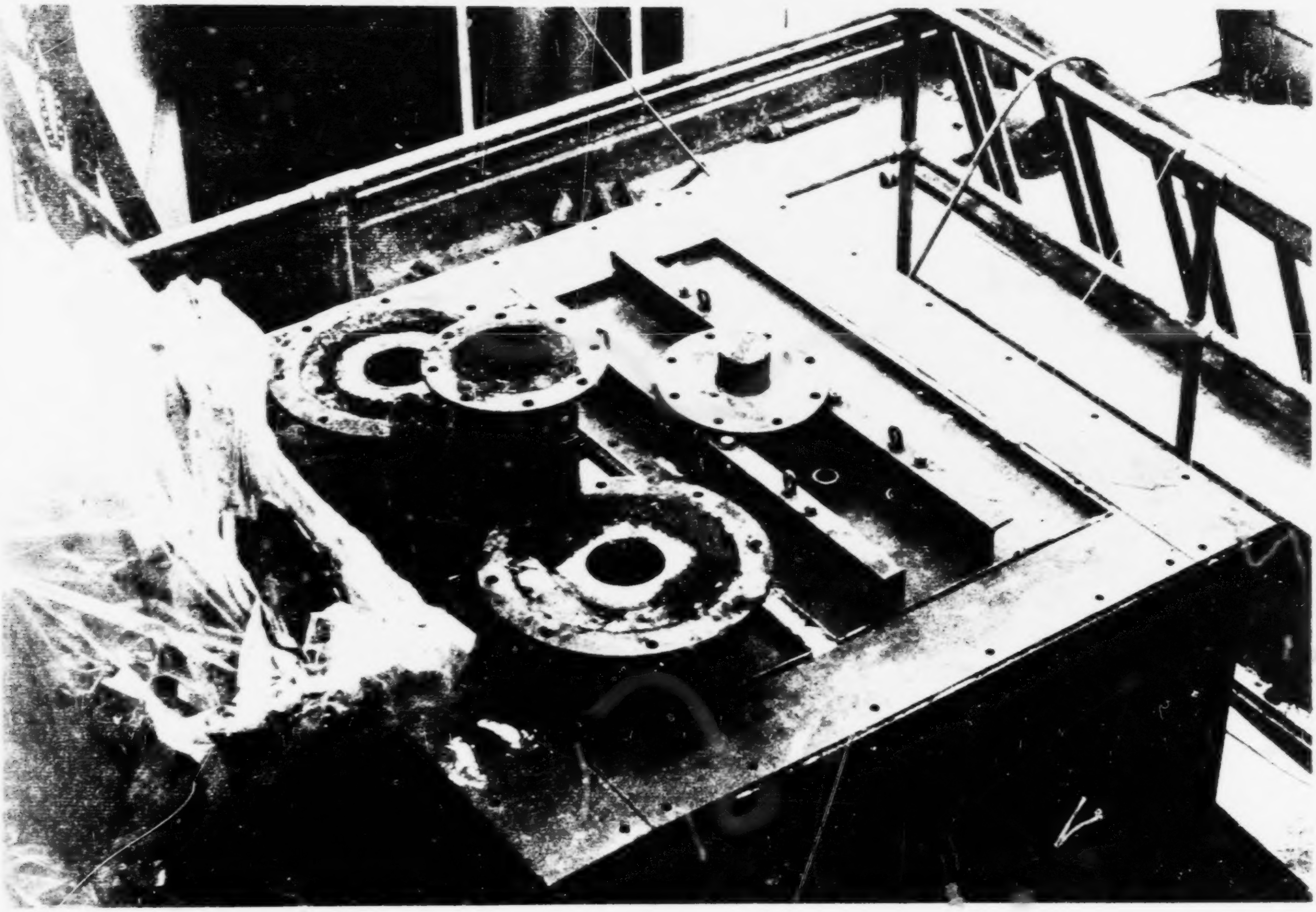

Fig. [illegible]

APPENDIX H

CAFB Pilot Plant Alarm Systems.

1. Alarm Actions

The installation is protected by a number of safety detection systems and there are four main actions that may be triggered off by an alarm depending upon which alarm is energised.

a. Alarm Action A

- Fire Valves Close on oil line flow and return at entry point into the building.
- Oil circulation pump stops.
- Gasifier control panel is alarmed - the consequences of this alarm may be controlled and are described below.
- Interior and exterior bells ring inside and outside laboratory.
- Red light comes up on auxiliary panel located on laboratory wall close to main door to 3A.

b. Alarm Action B

- Gasifier control panel is alarmed - consequences of this alarm may be controlled and are described below.

c. Alarm Action C

- Alarm light comes up on auxiliary panel.
- Interior and exterior bells ring inside and outside laboratory.

d. Alarm Action D

- Alarm light only comes up on main control panel.

e. Choice of main control panel action

There are various actions that may be preselected to follow an alarm signal to the main control panel.

- An internal bell can ring or may be made inoperative by a switch on the panel.
- Automatic shut down of all blowers, pumps, compressors except burner air supply which is manually controlled at all times. Alternatively this procedure can be made inoperative by a switch on the panel.

choice

- All alarm sensors may show as lights on the panel when alarmed.
- The first alarm sensor only will show as a light and any other alarms which are energised as a result of the automatic shut down will not show.
- The low hopper alarm has been arranged to show a warning light only and cannot be arranged to sound a bell or cause shut down.

Generally it is proposed to run the unit with the alarm bell in circuit and automatic plant shut down selected -this is necessary because failure of the gas pilot flame could result in unburnt gas forming in the boiler and ultimately an explosion.

f. Method of alarm display on main panel

The particular alarm light will be dependant upon the source of the alarm. Table H-1 lists the alarms and how they will be shown.

2. Alarm Systems

The installation is best considered as four main systems:

- The boiler and its cooling system
- The Gasifier
- The experimental burner on the boiler
- General alarms

a. The boiler and cooling system

The water in the boiler circuit is pressurised to 50 psi and pumped through a heat exchanger. The secondary side of the heat exchanger is pumped to an evaporative cooler located on the roof of the building. The following alarms are installed in this system.

(1) Failure of the cooler circulating pump will operate a differential pressure switch across the pump feed and delivery lines -

Result - Action (A) above

(2) High water temperature in the cooler water feed line - set to operate at 190°F.

Result Action A above

(3) High water temperature in the boiler - set to operate at 245°F.

Result Action B above

(4) Low Pressure in the pressurisation unit - set to operate at 43 psi.

Result - Action B above

b. Gasifier

The gasifier has many alarm circuits whose action has been described earlier. The following parameters are monitored:-

(1) High temperature in the gasifier bed - set to 950°C. Shown and alarmed from Guardian indicator on main panel.
(2) High temperature in regenerator bed - set to 1100°C. Shown and alarmed from Leeds and Northrup recorder.
(3) Gasifier distributor blocked - shown on switch and set to 15".
(4) Regenerator distributor blocked - shown on switch and set to 15".
(5) Regenerator bed low level - shown on switch and set to 10".
(6) Regenerator bed high level - shown on switch and set to 24".
(7) Down stream blockage indicated by pressure in gasifier gas space - shown on switch and set to 10".

All these alarms Result - Action B

(8) Hopper low level will cause a warning light only on panel.

Result - Action D

c. Experimental burner

The experimental burner will cause an alarm signal if there is a failure of:-

- pilot flame
- main flame
- low gas pressure set to 4" w.g.
- low air plenum pressure - set to 2" w.g.
- low pilot air pressure - set to .3 psi

Any one of these alarms will shown as alarm J on main panel.

Result - Action B

d. General Alarms

(1) Fire Detector

A fire detector is situated over the boiler and will alarm at 150°F.

Result - Action A

(2) Sump Level

In the event of the build up of liquid in the pit to about 1" deep over the floor of the pit an alarm will sound.

Result - Action C

(3) Emergency Stop Buttons

There are three red emergency stop buttons located

- Main door at 3A end of laboratory.
- Main door at main stores end of laboratory.
- Adjacent to ladder on side of the pit.

Result - Action A

(4) Emergency Stop Button on Main Control Panel

Button located at centre of main panel.

Result - complete unit shut down

3. Restart Procedure

In the event of an alarm showing it is important to determine the cause of the alarm and if it is safe to restart.

a. Restart after Action A

(1) Gasifier

The gasifier restarting procedure is described in the operating section, Appendix I.

(2) Auxiliary Panel

- If the audible alarm has not been muted it may be done by pushing the mute button on the auxiliary panel.
- The panel may be reset as soon as the alarm signal has ceased and the red light will go out.
- The two fire valves must be reset by lifting the lever and releasing the rewind mechanism by raising the rubber covered lever on the side of the panel and tightly winding up the string in a clockwise direction until the levers are set up as high as possible. The rubber covered levers can now be switched down.
  The oil circulating pump is restarted by depressing the green button on the side of the starter located near the pump.

b. Restart after Action B

Gasifier

The restarting procedure is described in Appendix I.

c. Restart after Action C

Check cause of high liquid level in pit and determine if it is safe to carry on.

d. Restart after Action D

Fill up hopper with limestone.

TABLE H-I

Summary of Pilot Plant Alarm System

		Indication	
	Source of Alarm	Auxiliary Panel	Main Control Panel
1.	Failure of water circulating pump.	Red Light	Red light titled "Auxiliary Panel"
2.	High water temperature on cooler feed line	Red Light and bells	Red light titled "Auxiliary Panel"
3.	High water temperature in the boiler	None	Red light titled "Boiler water or Burner"
4.	Low Pressure in pressurisation unit	None	Red light titled "Auxiliary panel"
		n.b. Red light shown on pressurisation panel, and its own bell rings.	
5.	Gasifier high temperature	None	Gasifier high temperature warning light.
6.	Regenerator high temperature	None	Regenerator high temperature warning light.
7.	Gasifier blocked distributor	None	Gasifier blocked distributor warning light.
8.	Regenerator blocked distributor	None	Regenerator blocked distributor warning light.
9.	Regenerator low bed level	None	Regenerator low bed level warning light.
10.	Regenerator high bed level	None	Regenerator high bed level warning light.
11.	Downstream blockage on gasifier outlet	None	Downstream blockage warning light.
12.	Low Hopper level	None	Low hopper level warning light.
13.	Experimental Burner Failure of: Pilot flame Main flame Low gas pressure Low air plenum pressure Low pilot air pressure	None	Red light titled Boiler water or Burner
14.	Fire Detector	Red light and bells	Red light titled Auxiliary Panel
15.	Sump Level	Red light and bells	Nothing
16.	Emergency Stop Buttons in building	Red light and bells	Red light titled Auxiliary Panel
17.	Emergency Stop Button on panel	None	None

APPENDIX I

CAFB PILOT PLANT OPERATING PROCEDURES

A. Preparation for Running

1 Operation of Services

a. Oil Ring Main

- Check tank contents.
- Check outflow temperature is about 140/150° F.
- Start compressor/oil pump on boiler by switching on at panel. Do not run for more than 5 seconds at first few runs in order to set thick oil moving. Continue until temperature gauge on outflow heater reaches 200/210° F. and then pump may be left on.
- The boiler heater circuit is now on automatic operation and is ready for use.

b. Kerosene

- Stored in 500 gal. tank - check that there is enough for anticipated usage. First check that isolating valve is turned off on pump supply feed pipe. Then open valve at barrel.

c. Nitrogen

- Stored in liquid N_2 tank. check that there is sufficient for anticipated usage.
- Check all valves are turned off at the bleed locations on gasifier before opening valve on manifold.

d. Propane

- Stored in 2 banks of cylinders outside building.
- Commence runs on the bank with least available so that replacement bottles may be ordered in reasonable time. Warn Purchasing at least 7 days prior to anticipated use and thereafter replacements can be obtained on 48 hours notice.
  Only turn on outside valve immediately prior to use for safety reasons.

e. Boiler and Cooling System

(1) Cooling System

- Open drain valve on cooling pump delivery which bleeds water to waste.
- Go up on roof and check valve is open on water feed to cooler - listen to check that water is flowing. If not - check all valves back to tower.
- Turn off valve on drain to conserve water until heating takes place.
- Warn Services of an anticipated soft water usage - maximum of 600 gph. Please give as much notice as possible.
- Turn on secondary side i.e. cooler pump - hold button in for at least 10 seconds to prevent a shutdown alarm.
- Turn on boiler circulating pump. No button to hold on this circuit.
- Turn on supply to cooler fan located on end wall (main stores end) n.b. fan is thermostatically controlled and will not cut in until pond water reaches 100°F.
- Check temperature setting of automatic mixing valve is set to 180°F and turn on electrical supply to valve. (Controller set on wall by pressurisation unit).

(2) Pressurisation System

- Check water level in storage tank is at rubber band marker and make up is turned on.
- Check nitrogen bottle pressure is above 500 psi.
- Turn on pressurisation system and check that pressure reaches 50 psi approximately. On start up bell will ring, cancel bell, then turn off N2 at cylinder.

(3) Boiler

- Check that main flue butterfly valve is at marked position. (Do not take any notice of "open" and "shut" marked positions as this was incorrectly fabricated).
- Check that water pressure in boiler is approximately 50 psi.
- Check that any sampling lines are correctly installed in flue.

2 Preparation for Running Gasifier

The gasifier is assumed to be cold and empty of bed.

a. General Check

- Carry out general inspection to ensure all main supply lines are complete and that no sampling or other instrumentation holes have been left open.
- Check condition of all drive belts.
- Check oil level in metering pumps, drain all air filters for moisture.
- Check all sampling pumps work.
- Check all analytical equipment and calibrate if necessary.

b. Control Unit

- Set alarm switch to "all alarms will show" position
- Set alarm bell to "mute"
- Set automatic shut down switch to "inoperative" position
- Turn on main power supply.
- "Burner or Boiler water" and "regenerator low bed" warning lights should come on. The former alarm is caused by pilot flame being out.
- Turn on regenerator blower and check that it will shut down if "automatic shut down" is selected. Reset alarm panel. Replace switch to "shutdown inoperative"
- Check that original two alarm lights show.
- Check and fill manometers if required.
- Start bleeds on injectors using direct N_2 bypass to avoid starting Compton compressors.
- Set pilot burner air supply to 6 cfm, turn on propane at external manifold, at plug cock on the wall, at valve over pit. Start main burner blower to pressurise plenum. Turn on pilot operating switch. The pilot should light and lock on. The pilot should stay alight and the warning light on the control panel go out.
- Set up recorders to 20 psi air supply. Switch in charts to check movements and recording pens are working. Turn off charts until required. Switch is in back of panel - plug and socket on each recorder.

B Warm up

The gasifier is heated over a 3 day period with a rate of temperature rise of about 12°C per hour.

It is important that heat be passed through the regenerator to make heat up of the refractory as uniform as possible. To secure a gas flow through the regenerator, the regenerator outlet valve must be open wide and the main boiler outlet tube damper closed. Be sure there is a nitrogen bleed through all injectors and pressure taps before starting.

The heat up sequence is as follows:

1. Start second stage air blower and line up 200 CFM air flow to gasifier. Remote valves set as follows -

1st Stage outlet	Closed
2nd Stage inlet	Open
Flue Gas Recycle	Closed

2. Turn on regenerator blower and set flow to about 3 CFM through regenerator.
3. Open air flow to fuel injector lines. Set to 5 CFM on each injector.
4. Start air flow to boiler main burners - Set to 900 CFM.
5. Turn on propane supply up to gasifier start up burner. Set ganged propane valve to "start" position.
6. Line up air flow to start up burner.
7. Turn on cooling water spray to gas burner casing.
8. Turn on electrical power to start up burner control and start burner.
9. Adjust propane air/fuel ratio to obtain about 3 times stoichiometric air to fuel.
10. Monitor temperatures and adjust main and regenerator dampers to obtain uniform temperature rise in gasifier and regenerator.
11. Adjust gas and air rate to burner to follow desired temperature schedule on TC 7 of the L & N Recorder (gas space below Lid).
12. Do not let air rate to distributor of gasifier fall below 100 CFM. When necessary, increase total air supply to gasifier. Remember that air to distributor is difference between total air and air to burner.
13. When gasifier temperature reaches 700°C begin kerosene firing. Do not start Kero firing without technical man on unit with operator.
14. Kerosene should be cut in very slowly at first, one pump at a time, to avoid upset to temperatures. Once kero feed is started, gradually raise kero rate to follow temperature schedule to 850°C. Air rate to plenum may have to be increased to provide sufficient air for kero combustion. There should be at least 30 CFM air through the plenum for each gallon/hr of kerosene feed.

15. When gasifier temperature reaches 850°C begin reducing gas and air to burner while increasing kero and air to gasifier to maintain temperature. Gas rate should be reduced to about 100 CFM, burner air to about 60 CFM, and total air about 350 CFM.

16. Fill limestone hopper while heat up is in progress.

17. Start air flow at 5 CFM to limestone feed system.

18. Start limestone feed vibrator at low setting (about 1) and begin adding stone to unit. Do not start limestone addition without technical man present.

19. Observe temperatures and fluidisation as soon as limestone addition starts to be sure that stone is being heated and fluidised throughout. If necessary adjust air and firing rates to obtain uniform heat up. Avoid letting stone temperature fall below 800°C.

20. Gradually add limestone to reach specified bed level on manometer. Do not exceed specified vibrator setting while adding bed, or fill line will block.

21. Start Compton compressors, raise regenerator air flow to 15 CFM, and begin bed circulation between gasifier and regenerator as soon as bed depth reaches 10 in. w.g.

22. When bed level reaches 12 in.w.g. the start up gas burner can be turned off. Start a 4 CFM air flow through burner cooling inlet to keep burner head cool and block main air line to burner.

C. Initiating Gasification

When the gasifier is at operating temperature with a desired bed level established and circulating, gasification can be started. The procedure used is to switch from kerosene combustion to fuel oil combustion, increase oil rate to the stoichiometric ratio to eliminate oxygen from the system, and then increase oil rate to that required for gasification.

1. Starting point is with gasifier at about 850°C with at least 12 in.w.g. bed depth, and good solids circulation in both directions between gasifier and regenerator. Temperature is maintained by kerosene combustion in the bed with excess air.

2. Increase air flow to main burner to 1100 CFM total of which 150 CFM is passed through the premix nozzle.

3. Check that pilot flame is strong and stable. Do not proceed until it is.

4. Check cooling system to be sure it is operating satisfactorily.

5. Line up flue gas recycle through second stage blower and adjust total flow to gasifier plenum to 260 CFM; 30 CFM flue gas recycle and 230 CFM fresh air.
(Procedure was changed during 3rd pilot plant run to pass flue gas recycle through both first and second stage blowers).

6. Stop oil pumps and switch pump suction lines to fuel oil supply from kerosene supply.

7. Set centre pump to deliver 78 lb/hr fuel oil (stoichiometric) and start pump.

8. Set pumps 1 and 3 to deliver 130 lb/hr fuel oil each, and when gasifier bed temperature reaches 900°C, switch on both pumps. Normally pumps 1 and 3 will be required about 15 seconds after starting pump 2 at stoichiometric oil rate.

9. Main boiler flame should ignite within about 15 seconds of starting pumps 1 and 3.

10. Raise pump 2 setting to deliver 130 lb/hr fuel oil.

11. Adjust flue gas recycle rate to control temperature at desired level.

12. Check that cooling tower fan starts when secondary water temperature reaches the set point.

13. Begin limestone feed addition.

14. Bring all conditions to running specifications.

D. Solids Handling

1. BED REPLACEMENT

It is necessary to add fresh bed to the gasifier to maintain efficiency of operation. Bed material will be drawn out of the regenerator at 18/20 lbs/hr using a vibratory feeder to an external storage hopper on the floor of the pit. The gasifier bed level will then drop and start a vibratory feeder which will add material from a storage hopper mounted on the gasifier lid. The storage hopper has a low level warning and may be replenished from a ground level supply of raw material by a suction lift pipe.

It is most important that gas cannot enter into the hopper and deposit tar which would stop the feed of material. The feed system is therefore a sealed system valved in three places and with an inert gas purge in the feed pipe to discourage back flow. During hopper refilling the vibratory feeder must be turned off and the system must be isolated by the air operated valve above the gasifier lid before opening the valve on the material feed line and the valve on the vacuum blower. Refilling will be carried out when the low level alarm shows a nearly empty hopper and the requisite amount of material as shown by the full line on the storage hopper may be added without overfilling the hopper. Immediately after filling close the valves on the feed line and vacuum blower and then open air operated valve and turn on the switch on the vibrator feed.

Summary of operating procedure for gasifier hopper filling:

F.H. is feed hopper located above gasifier.
S.H. is ground level storage hopper.

a. Low level alarm shows on console.
b. Turn off vibratory feeder switch and wait for line through valve to clear.
c. Turn off isolating valve by turning 3 way air cock to closed position.
d. Open upper valve on vacuum blower - check lower valve is closed.
e. Open valve on feed pipe near S.H.
f. Check S.H. level is at correct height.
g. Switch on vacuum blower until all material has gone into F.H. Low alarm should now go out.
h. Shut valves on vacuum blower and feed line.
i. Open air operated isolating valve.
j. Turn on vibrator switch.
k. Refill S.H. to correct level for next filling operation.

2. BED REMOVAL

Hot bed material will be removed from the regenerator by manually operating the vibratory feeder which will feed the material into a storage hopper - trials will have to be made at temperature to determine if a low steady draw off can be achieved or if it will be necessary to draw off at intervals. A nitrogen purge is available if required.

3. DRAINING REGENERATOR CYCLONE

This cyclone may be drained periodically as shown to be necessary by experience. Normally the upper valve is open and hopper drain valve closed. To empty hopper just shut upper valve and open lower valve, tap the upper sides to free any fines adhering to the walls. Close lower valve and open upper valve when emptying is completed.

4. GASIFIER BED SAMPLING

Samples may be taken from either of the gasifier drains by closing lower valve, opening upper valve and bleeding in nitrogen for a brief period. Then open lower valve with a suitable container in place and collect sufficient material.

E. Carbon Burn-out Procedure

The carbon-burnout procedure is initiated when the gasifier pressure above the bed reaches 20 in.w.g. or when pressure drop through the cyclone and duct system begins a rapid rise.

1. Preparations

a. Call out additional assistance if needed.

b. Connect burn-out gas supply fitting to gasifier lid.

c. Prepare gas sampling lines from plenum gas feed and lid gas feed lines to O_2 and CO_2 meters.

d. Shut off any air bleeds through lid

2. Bed Sulphation

a. If necessary, lower bed level by draining gasifier solids.

b. Reduce gasifier temperature to below 850 °C by increasing flue gas recycle rate.

c. Line up steam to steel cyclone gas outlet tubes.

d. Switch fuel injector bleed gas from air to nitrogen at 2 CFM each.

e. Shut off fuel to gasifier by stopping pumps and closing valves in fuel lines.

f. When main flame goes out, close air valves to burner and close flue outlet damper. Close fresh air inlet to gasifier blower.

g. Set flue gas recycle rate to maximum.

h. Set high nitrogen flow rate to lid. This flow can be stopped later.

i. Control bed temperature to 900-950°C range by regulating flow of fresh air to main burner. This air circulates back through flue gas recycle line to provide oxygen for bed sulphation. Proper oxygen concentration is about 10% O_2 in gas to plenum.

j. When sulphation is complete, temperatures down stream of bed will begin rapid rise and bed temperature will begin to fall. Sulphation normally requires about twenty minutes. At this point close air flow valve to gasifier distributor and reduce regenerator air flow to about 3 CFM.

3. Carbon Burn Out

   a. Start flow of flue gas and air through gasifier lid and observe effect on gas space and cyclone temperatures.

   b. Adjust fresh air flow to main burner to obtain desired O_2 content of gas to gasifier lid.

   c. Adjust steam rate to cyclone tubes to avoid over-heating. Hold cyclone tube temperature at 900°C or lower.

   d. Gradually increase gas flow through lid to about 50 CFM. Continue close control of temperatures.

   e. When carbon burning at cyclone tubes is complete, gas flow and O_2 content can be increased to complete burning down stream of cyclones.

   f. Burn out is complete when all duct temperatures begin falling. Care must be taken not to allow bed temperature to fall too far during burn-out or difficulty will be experienced during restart.

4. Restart

   a. Stop air flow to gasifier lid.

   b. Close steam valves to cyclone tubes.

   c. Open main flue dampers.

   d. Open air flow to main burners.

   c. Start pilot light.

   f. Partly close flue gas recycle valve and open air suction valve to gasifier blower.

   g. Set gas flow to gasifier plenum to 260 CFM.

   h. Return fuel injector gas to air from nitrogen.

   i. Start fuel oil injection at low rate to return gasifier temperature to 850°C.

   j. Initiate gasification by usual procedure except that no switch over from kerosene is required.

APPENDIX J

Operation of Continuous Pilot Plant

This appendix contains a summary of operations during the three 1971 runs of the CAFB pilot plant, data plots and tables, photographs and a description of the unit at the end of the third run.

A. RUN 1

The continuous CAFB pilot plant was operated as a gasifier for 160 hours in August. During this period, a number of major process features including continuous sulphur removal, production of a highly combustible gas giving a clean luminous flame, gasifier and regenerator temperature control, production of a concentrated SO_2 stream from the regenerator, and continuous solids circulation were demonstrated.

However, several operating difficulties were identified which prevented full achievement of the 200 hour run objective and which caused several interruptions in gasification. To a large extent these difficulties were of a mechanical nature which slight modifications remedied. Two process related problems were also identified which required modified operating procedures for their solution.

1. Description of Operations

a. August 1 - 5

The CAFB gasifier unit was preheated at the end of July, and a bed of lime was established. Hot lime circulation tests were conducted during the period of August 1 - 5. It was established that best circulation rates could be maintained if pressure differential between gasifier and regenerator was kept to below 5 inches w.g. A valve was installed in the regenerator outlet line to facilitate control of this pressure difference. Solids transfer rates of over 800 lb/hr were reproducibly demonstrated during this period.

A test of the fuel feed system revealed that the shrouded injector lines would give an unacceptably high pressure drop at start-up (when the final section of injector line was cold). The fuel system was modified by removing the inner injector tubes and allowing the fuel flow to pass through the larger diameter tubes formerly used as shrouds. A stream of air injected with the fuel keeps the velocity high and prevents coking in the inlet tube.

b. August 6 - 9

After making additional tests of the pilot flame system to verify conditions which assured its stability under hot operating conditions, gasification trials were initiated on August 6th. Two tests were made with the back boiler door open and an extension to the fire tube in place. Only short runs could be made with this arrangement, since no flue gas recycle was available, and ability to control temperature was limited. The procedure used in changing from combusting to gasification conditions was to:-

(1) raise fuel rate to the stoichiometric level for approximately ten seconds to sweep air from the gasifier with combustion products;

(2) raise fuel rate to 500 percent of stoichiometric to initiate gasification.

Smooth ignition of the main flame took place at once when gasification started.

The first open tube gasification trial lasted seven minutes, the second approximately fifteen minutes. Both tests were carried out long enough to observe that the main burner flame was stable and that the start-up procedure appeared to be safe. Both tests were terminated by turning off fuel oil flow and returning to kerosene combustion when gasifier temperature exceeded 920°C. During these brief tests it was noted that regeneration had started as regenerator temperature began rising. For the second test it was noted that sufficient sulphur had been deposited in the gasifier bed to cause a sharp temperature rise when fuel flow was discontinued.

The extension tube was removed and the boiler door closed after the second open tube test. A gasification trial was then started with the complete boiler, flue gas stack, and flue gas recycle system in operation.

The first continuous system trial lasted apprximately 12 hours. It was terminated when a gradual increase of back pressure on the gasifier was accompanied by a loss of air flow rate to the gasifier plenum. Gasifier pressure had increased from an initial level of 8 inches to 14 inches w.g. when the shutdown was made. Reduced capacity of the air blower was later traced to belt slippage. Pressure rise in the gasifier was attributed to build up of carbon between gasifier and outlet duct. The obstruction causing this pressure rise was removed by returning the unit to kerosene combustion conditions with excess air. The excess air apparently burned out any carbon accumulation that had existed.

During the twelve hour operating period, good temperature control of gasifier was maintained by adjusting relative flue gas and air rates, and regenerator temperature was controlled by adjusting solids circulation rate. However, SO_2 concentrations from the regenerator never exceeded 4.3% by vol. Good solids circulation rate was maintained right up to the shut-down.

For a period following the shut-down of gasification, all air and fuel flow to both gasifier and regenerator beds was stopped except for small air bleeds through the injectors and pressure taps, and a slight bleed to keep the regenerator distributor holes free of fines. When air flow was resumed, an obstruction was observed in the regenerator bed itself and in the regenerator-to-gasifier transfer line. The period from August 7th to August 9th was spent in removing these obstructions and in re-establishing solids levels, circulation rate, and temperature under full combustion conditions.

c. August 9 - 11

Gasification resumed on August 9th at 15.30 hours and continued until 10.00 on August 11th. Gasification was interrupted because of two operating difficulties: (1) pressure in gasifier had risen to 14 inches w.g. and flue gas recycle rate was limited by reduced blower capacity at the higher back pressure; (2) regenerator instrumentation indicated an obstruction in regenerator outlet. This was subsequently traced to a plugged pressure tapping which caused a false pressure reading.

When gasification stopped, a flow of fluidising air was maintained in the regenerator. The gasifier bed was slumped and then pulsed with short bursts of fludising air over a period of three hours to convert the bed sulphide to sulphate without excessive temperature rise. The bed was then put on full combustion with excess air to burn out any carbonacious obstructions. Before resuming gasification it was necessary to remove an obstruction that had formed in the gasifier-to-regenerator transfer line. During the August 9 - 11 operating period conditions were varied to increase regenerator SO_2 concentration and to improve sulphur absorption in the gasifier. Desulphurisation of over 90% was achieved briefly, and SO_2 concentrations over 10% were reached. During this period it was apparent that the gasifier cyclones were functioning as large quantities of fines were removed from the regenerator cyclone drain.

d. August 11 - 15

Fuel oil was introduced to resume gasification at 22.40 hours on August 11th. Continuous gasification was maintained until 13.30 on August 15th when gasifier internal pressure had increased to 24 in. w.g. This period of approximately 87 hours provided the most consistent demonstration of gasification, desulphurisation, and regeneration in the first test. Conditions maintained in this period are shown as a series of log plots in Figures J-1 through J-7.

Desulphurisation efficiency varied from 75% to 97% and averaged 90%. The SO_2 concentration in the regenerator off gas averaged 7.5% for the entire period and exceeded 10% a number of times.

Other average conditions for the period are listed below in Table J-I. It was evident that something had happened to change performance of the gasifier cyclones between the end of the August 9 - 11 period and beginning of the August 11 - 15 period, as practically no fines were recovered from the regenerator cyclone in the latter period. Fines evidently were lost directly from the gasifier into and through the boiler.

Table J-I

CAFB Pilot Plant Average Run Conditions
11 - 15 August, 1971

Gasifier	
Temperature	869°C
Bed Depth	15.6 in. w.g.
Bed Density	49.5 lb/ft^3
Superficial Gas Velocity(a)	3.2 ft/sec.
Oil Feed Rate	361.4 lb/hr
Stone Feed Rate	43.8 lb/hr
Calcium/Sulphur Ratio	1.66 mole/mole
Air/Fuel Ratio	16.8% of Stoichiometric
Sulphur Removal Efficiency	90%
Regenerator	
Temperature	1047°C
Air Rate to Distributor	16.7 cu.ft./min.
SO_2 in off Gas	7.5%
O_2 in off Gas	0.47%
CO_2 in off Gas	2.75%
Solids Discharge Rate	10.2 lb/hr.

(a) velocity of air and flue gas fed to plenum taken at gasifier bed temperature and top bed empty cross section area of 5.2 sq.ft.

e. August 15 - 18

After termination of gasification on August 15th, the gasifier was pulsed with air to sulphate the bed without excess temperature rise. Kerosene combustion with excess air was then resumed to decoke the gas outlet. Following this operation it was found that partial obstructions had developed in the solids transfer lines, and efforts to remove them were unsuccessful. Gasification with limited circulation was conducted from 2200 hours on 17 August until circulation stopped at 1900 hours on 18 August. Gasification without circulation was continued until 01.00 on 19 August when fuel was stopped because of gasifier pressure increase to 20 in. w.g. At that point a controlled bed sulphation was conducted, but air flow was stopped before burning off carbon above the bed in order to preserve any obstructions which had formed. The unit was then cooled and opened for inspection.

2. Condition of Unit After Run

The unit was inspected for refractory cracks and was found to have no new cracks since the initial firing-in period. However, some of the old cracks in the lower portion of the unit appeared to have increased in size. These enlarged cracks were in the bottom lift of concrete which was the coldest portion of the unit. None communicated between gasifier and regenerator. Some led out to the space between concrete and insulation. There appeared to be no cracks in the transfer lines or in the cyclone walls. The cyclone walls did show appreciable roughening.

Accumulations of solids were observed at several locations in the pilot plant when the unit was opened. Samples of these accumulations were analysed, and their chemical composition is reported in Table J-II. Their location and general appearance is described in the following section.

a. Gasifier

Walls of the gasifier above the bed were coated with a very thin layer of carbon. There appeared to be little carbon on the walls below the bed level. There was a carbon-free area in the gasifier lid around the limestone feed point. Apparently the air bleed in the limestone inlet had prevented carbon formation at that point. The gasifier bed wall on the fuel injector side showed a roughened appearance both below and above the bed level. This contrasted with the smooth wall on the opposite side.

Mounds of loosely agglomerated lime particles were found beneath and surrounding each of the fuel injection nozzles. These mounds were removed from the gasifier intact, but in several days they lost strength and disintegrated into lime particles. They appeared gray and relatively free of carbon. A similar, but much smaller mound was found just beyond the outlet tube where regenerator solids return to the gasifier bed.

b. Gasifier Outlet Ducts

The thin layer of carbon on the gasifier walls thickened in the region of the outlet ducts leading to the cyclones. Formations appeared to have grown across the duct inlets obstructing a large portion of their areas. These deposits extended only a short distance into the ducts and appeared to have formed in the eddies of the entrance region. These deposits are the probable cause of much of the increase in pressure drop between gasifier and cyclones during gasification. The deposits were hard and dense. They appear to consist mostly of carbon but contain some lime fines.

c. Cyclone Outlet Tubes

Both cyclone outlet tubes, made of type 310 stainless steel, had been burned away, and only short stubs remained. Portions of the tubes had fallen into the cyclone bodies blocking the drain lines. Much of the cyclone bodies were filled with loose lime particles.

d. Regenerator Bed

Walls of the regenerator in the bed space were clean. However, two large solid lumps occupied the lower portion of the bed just above the distributor. Each lump obstructed a portion of the bed cross section, but a passage remained between them. The material from which these lumps were formed was quite hard and dense.

The regenerator walls above the bed were clean for a considerable distance above the bed. However, a rough surfaced growth had formed just beneath the outlet duct at the regenerator top. This formation did not reduce the opening below the outlet duct diameter, but it did form a conical transition shape from the bed diameter to the duct diameter.

e. Solids Transfer Lines

The catch pocket in the gasifier wall was crusted over with a skin-like layer of agglomerated solids. Otherwise the gasifier-to-regenerator transfer line was clear. On the regenerator-to-gasifier side, a lump of hard solid was firmly attached to the wall of the mixing pocket just above the agitator air inlet. Another hard obstruction was located at the bottom of the solids down-comer where it entered the mixing pocket.

f. Gas Transfer Duct

The major portion of the gas transfer duct was clear except for a thin deposit of carbon. However, a growth of carbon had formed at the junction where gas streams from the two cyclones combined. The growth acted as a fairing in the dead zone between branches of the Y. Other small growths protruded into the ducts above the cyclones where the gas changed direction from vertical to horizontal. Again the growth filled what appeared to have been the down stream eddy spaces.

g. Burner

The steel inner liner of the burner up-stream of the main air mixing point was badly burned on its lower surface but not on its top. It appears that carbon particles may have accumulated in the bottom during decoking operations and then burned at a high temperature. The burner face and the stainless steel baffles installed for pilot flame stabilisation were intact and in good condition.

h. Boiler

The boiler end space and lower tubes contained a considerable accumulation of lime fines. Approximately 1300 pounds of solids were recovered from the boiler and its tubes. The accumulations were all loose and probably could have been removed during operation had there been appropriate drain points.

TABLE J-II

Composition of Solid Residues from Pilot Plant after First Run

Sample Location	Raw Stone	Regenerator Bottom Deposit	Regenerator Cyclone Fines	Gasifer Cyclone Deposit		Boiler Tube Deposit	
				Inlet	Outlet	1st Pass	3rd Pass
Composition							
CaO wt.%	53.8	39.5	80	24.5	70.2	70.1	74.8
MgO "	0.5	0.6	0.6	0.6	0.6	0.6	0.6
S_1O_2 "	0.3	0.3	0.5	7	0.7	0.7	0.7
Na_2O ppm	34	740	195	540	145	210	220
Fe_2O_3 wt.%	0.2	0.39	0.49	0.89	0.94	0.94	0.61
Al_2O_3 "	0.2	0.2	0.2	5.9	0.4	0.3	0.3
Carbon "	-	< .1	0.25	46.4	16.9	.25	1.18
Total Sulphur wt.%	-	23.17	8.62	2.87	8.14	3.7	5.08
Sulphide Sulphur wt.%	-	0.07	7.26	2.39	7.90	1.53	3.52
Vanadium wt.%	-	0.08	0.13	0.23	0.37	0.4	0.3

3. Mechanical Trouble Areas

Three operational difficulties of a primarily mechanical nature contributed to run interruptions and non uniformity of operating conditions. These were: (1) Slippage of drive belt on gasifier air blower;
(2) Blockage of limestone addition line.
(3) Plugging of flue gas compressor suction line.

Slippage of the drive belt on the gasifier blower was a purely mechanical problem easily corrected when recognised. However this slippage decreased blower capacity during early stages of the gasification run and forced premature shutdown after moderate gasifier pressure rise on August 7 and August 11. Tightening the drive and application of a belt dressing eliminated the problem after August 12.

Periodic plugging of the limestone addition system was a more persistent problem that never was completely eliminated during the run. In early stages of the run it was aggravated by use of flue gas recycle from a Nash compressor as bleed gas. This flue gas was wet and caused the stone to become sticky and prone to plug the line. Replacement of the flue gas with air and installation of a clearing rod with block valve and stuffing box greatly improved the operability of the system but did not eliminate occasional stoppage, especially when addition at a low rate was desired. A larger stone addition line and a greater air bleed rate was provided for the next test to improve reliability of stone addition. The stone flow rate interacts with gasifier temperature control, since stone addition imposes a substantial heat load on the system. Improving stone addition uniformity eases the task of gasifier temperature control.

Periodic plugging of the suction line to the flue gas compressor eventually led to abandonment of the use of flue gas for instrument bleeds and injection gas to the solids transfer system. Replacement of this flue gas with air eventually created additional problems. The plugging agent in the flue gas recycle line was lime fines which built up at the point in the line at which temperature fell below the dew point of the flue gas. The flue gas compressor suction had been equipped with a water scrubber to knockout the lime fines. However, the scrubbing water cooled the inlet line enough to cause condensation of water vapour at the inlet to the scrubbing section. This inlet then quickly plugged. Although a more sophisticated scrubbing system with filters and preheat could be provided, it was decided to use nitrogen from the ERCA site supply as the gas for bleeds and solids transfer injectors.

4. Process Operating Problems

Two process features have the potential for creating operational difficulties, but results of the first continuous pilot plant test indicated that control of these features could be achieved. These process problem areas were:(1) Formation of deposits of carbon and lime fines in critical operating points between gasifier and boiler, (2) Agglomeration of sulphided lime particles when regenerated in a defluidised state.

5. Process Variable Relationships

There was no attempt in the first run to carry out a programme of variable studies. The objective was to hold most variables at relatively constant levels, and this precluded much possibility of variable correlations. However, experience in operating the unit demonstrated interdependence of a number of factors.

a. Limestone Feed Rate

Limestone feed rate influenced most aspects of gasifier operation, and it was unfortunate that this feed rate often was erratic. With a constant limestone addition rate, the unit was self stabilising and tended to line out at a constant set of conditions. However, sudden changes in feed rate caused variations which propagated throughout the system. Increasing limestone feed rate increases heat demand of the gasifier both to heat the stone and to achieve calcination. To provide additional heat, the air/flue gas ratio must be increased or gasifier temperature will fall. Increasing the quantity of fuel combusted, together with the CO_2 released by calcination, increases the gas velocity from the gasifier. This in turn increases the pressure drop through the gas ducts and cyclones and raises gasifier pressure. This change in gasifier pressure relative to regenerator pressure affects solids circulation rate between the beds and in turn affects regenerator temperature and SO_2 evolution rate. An interruption to bed addition allows gasifier bed depth to fall. Finally, a prolonged change in limestone addition rate affects activity of the bed for sulphur recovery and changes sulphur removal efficiency.

b. Sulphur Removal Efficiency

Sulphur removal efficiency varied from over 98% near the beginning of the test period to a low of 75% near the end, with fluctuations between 80 and 98% throughout the test. The major cause of fluctuations is believed to be variations in limestone addition rate. However, there is not yet a clear correlation due to variation of other factors such as sulphur content of the circulating lime. It is evident that sulphur removal efficiency of over 97% was obtained with about 1.5 to 1.7 mol Ca/mol S lime addition rate. Following an interruption to bed addition early on 15 August, bed depth decreased, and efficiency fell to about 75%.

c. Gasifier Temperature and Air/Fuel Ratio

Gasifier temperature was self-stabilising so long as fuel rate, air rate, and limestone addition rate were constant. There was no automatic control device for temperature, so any changes required were made by the operator. The principal cause of the temperature variations that did occur was interruptions to limestone addition which required adjustment of the ratio of flue gas recycle to fresh air addition.

The average air/fuel ratio was 17% of stoichiometric which gave a rich, highly combustible fuel gas. Even lower air/fuel ratios, down to 15% of stoichiometric, were used during periods of no limestone addition. Substantial cooling by flue gas recycle was being used even at these low air fuel ratios which means that, from a heat balance standpoint, even lower air/fuel ratios are possible.

TABLE J-III

Composition of Regenerator Samples
Removed from Pilot Plant during Run 1

Sample	Date-Hour	12-8-0130	12-8-0730	12-8-1430	12-8-1730	12-8-2030	13-8-0430	13-8-0730[a]			14-8-1030[a]			15-8-0030	15-8-0830
Composition								Total	Brown	White	Total	Black	White		
CaO	wt.%	87.7	-	-	-	-	-	89.7	79.4	88.0	94.1	90.5	82.7	-	-
MgO	"	0.6	-	-	-	-	-	0.6	1.1	1.1	0.6	0.6	0.6	-	-
S_1O_2	"	1.0	-	-	-	-	-	0.8	2.5	0.2	0.9	0.7	0.3	-	-
Na_2O	ppm	185	-	-	-	-	-	195	525	42	295	420	51	-	-
Fe_2O_3	wt.%	0.9	-	-	-	-	-	0.69	1.05	.16	0.3	0.36	.11	-	-
Al_2O_3	"	0.4	-	-	-	-	-	0.4	1.8	0.2	0.4	0.4	<0.2	-	-
Carbon	"	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1
Total Sulphur	wt.%	3.37	4.36	4.91	3.13	3.33	3.15	2.38	2.61	2.35	5.10	8.52	5.36	4.38	4.73
Sulphide Sulphur	wt.%	2.20	3.45	3.73	0.73	2.04	0.84	0.96	2.05	0.90	4.44	8.15	4.41	3.20	4.57
Vanadium	wt.%	0.75	0.68	0.60	0.68	0.67	0.77	0.73	0.91	0.18	0.8	0.65	0.09	0.77	1.02

(a) Sample subdivided by colour of particles

d. Regenerator SO_2 Concentration

This test provided the first data on CaS regeneration under continuous operating conditions. The results show that continuous regeneration is feasible and that high (of the order of 10%) SO_2 concentrations can be obtained below 1100°C. The SO_2 concentrations achieved averaged about 50% of the equilibrium values for the corresponding temperatures as shown in Figure J-7.

Composition of regenerator solids samples removed during the run are listed in Table J-III.

e. Sulphur Balance

The following sulphur balance was obtained during the period of August 11 - 15 in the pilot plant:-

Sulphur Input 361.4 lb/hr. fuel x .0248 fr S = 8.97 lb/hr
Sulphur Out in fuel gas 8.97 x 0.1 = 0.90 lb/hr
Sulphur Out Regenerator as SO_2 = 6.82 lb/hr
Sulphur on Solids from Regenerator = 0.4 lb/hr
Sulphur on Fines Passing to Boiler = 0.64 lb/hr
Unaccounted for Sulphur = 0.21 lb/hr
Sulphur Balance 100 x 8.76/8.97 = 97.5%

6. Unit Modifications

The following modifications were made to the pilot plant between Runs 1 and 2.

a. A nitrogen supply was installed for gas bleeds to avoid need for flue gas compression and to provide an inert gas source when the boiler was not operating.

b. The limestone feed line was increased in diameter with provision for additional air bleed.

c. Entries to the gasifier outlet ducts were rounded to provide smooth entrance regions.

d. View ports were installed in the lid, and provision was made for directed air injection to burn carbon at the outlet ducts.

e. Cyclone outlet tubes were removed.

f. The burner nozzle shape was modified to eliminate accumulation of loose carbon.

g. Provision was made for draining lime fines from the boiler end during operation.

h. Fuel injector nozzles were reduced in diameter.

B. RUN 2

Heating of the CAFB pilot plant for its second test began on October 1, 1971. Gasification started on October 7 and terminated on October 11 because of excessive lime losses from the unit.

Most of the mechanical problems which affected operations in August were absent in October. These items included blower belt slippage, plugging of lime transfer lines, and blockage of the limestone feed line. However, a high rate of lime loss from the gasifier interacted with many features of the process to reduce unit performance and prevent obtaining information in the region of principal interest.

Desulphurisation was essentially complete throughout the test period, as measured by SO_2 content of the boiler outlet gas. However, much of the sulphur was absorbed or reacted with lime that passed through the boiler rather than through the regenerator. Also, limestone replacement rate was higher than would be desirable for a commercial plant.

The range of operating conditions is shown in a series of log plots in Figures J-8 through J-15.

1. Description of Operations

Temperature of the gasifier interior was raised to 900°C by firing with propane and kerosene between October 1 and October 4.

Lime addition, using material removed from the bed after the August run, started on October 4. By October 5, a bed level of 15 inches w.g. had been established and bed circulation was started. Circulation was uneven however, and attempts to find a set of conditions giving steady circulation were unsuccessful. The difficulty was traced to a crack which had developed in the regenerator distributor. This distributor was replaced on October 6. However, under hot conditions it was not possible to obtain a perfect seal between the regenerator base and distributor with the result that some air bypassed the distributor holes. Good fluidisation and solid circulation were obtained when gas flow to the unit was resumed. Circulation calibrations and final unit checks were completed and gasification initiated at 21.14 hours on October 7.

There were no interruptions to gasification for the 91 hour period of operation through 16.30 on 11 October. Sulphur removal from the flue gas was essentially complete for the entire period. However, lime losses, and the consequent limestone make-up rate were very high, and less than half of the sulphur was being removed as SO_2 in the regenerator off-gas. The remainder was absorbed on lime which either was removed from drain points on the boiler or was lost as fines with the flue gas.

Attempts to reduce lime losses to a reasonable level by variation of operating conditions met little success, and therefore the run was stopped at the end of 91 hours.

2. Lime Losses

Figure J-8 is a time plot of average limestone feed and lime removal rates for various periods of the run. Since lime losses were higher than the target lime replacement rate, no lime was removed from the regenerator other than periodic analytical samples. As Figure J-3 shows, the rate of lime loss varied from about 20 to 50 lb/hr during the run. Limestone replacement rates as high as 135 lb/hr therefore were needed to maintain gasifier bed level. In the early portion of the run, bed level had been allowed to fall from an initial level of 18 inches w.g. In the mid portion of the run, the level was stabilised at 13 to 14 in. w.g. while temperature was maintained at 870-900°C. This condition gave the highest fines losses.

In the final portion of the test, bed level was allowed to fall to 11 inches w.g. and gasifier temperature was lowered to below 850°C. These changes reduced the fines losses, but not enough to give satisfactory performance. As Figure J-9 shows, the calcium oxide/sulphur ratio varied from about 2.5 to 4.6 mole/mole.

The major cause of the high lime fines losses during this run was elimination of the central cyclone outlet tubes from the unit.

However, a contributing factor was the use of high velocity air jets in the fuel injectors. After Run 1, a 3/16" diameter nozzle had been inserted in each fuel injector to improve dispersion of the oil. During start-up it was observed that air velocities of about 10 CFM injected with the fuel gave improved combustion in the bed and these high velocities were maintained through the gasification run. These conditions produced sonic jet velocities at the fuel inlet and evidently produced serious lime attrition.

The method of adding limestone directly through the gasifier lid also contributed to the losses as some of the stone was blown through the outlet ducts before it even reached the bed.

3. Gasifier Outlet Pressure Drop

The first run encountered a gradual rise in pressure drop between the gasifier and the transfer duct. Examination of the unit at the end of that test indicated that most of the resistance had been due to an accumulation of lime and carbon at the entrance to the gasifier outlet ducts leading to the cyclones. In the October run, the rate of pressure rise was much lower, and for over 70 hours appeared to be negligible. However, for the final 20 hours, the pressure did rise at a significant rate (Figure J-10).

4. Regenerator Operation

The high lime loss rate in the October run caused much of the sulphur absorbed in the gasifier to be removed with the solids that passed into the boiler. This factor reduced the sulphur content of lime going to the regenerator and reduced regenerator efficiency.

Consequently, the sulphur dioxide content of the regenerator off gas remained below 4% during much of the run and failed to exceed 8% at all. The selectivity of regenerating sulphide to oxide also was lower than in the first run, because the quantity of sulphide oxidised to sulphate in heating the solids to regenerator temperature represented most of the sulphide available. Figure J-15 shows the log of regenerator SO_2 output (as equivalent sulphur lb/hr) and of CaS to CaO oxidation selectivity. This selectivity generally was below 50% compared with values from 50 to 80% obtained in the first run when the lime entering the regenerator contained more sulphur.

Figure J-14 shows the oxygen and carbon dioxide concentrations in the regenerator gas. Both were higher than experienced in the first run due to the lower sulphur on lime. The very high oxygen level observed early on October 11 was obtained during a test in which solids circulation was reduced to a very low rate. This test indicates some passage of gasifier product through the transfer line into the regenerator since there could not have been enough carbon on the slowly circulating lime to account for the significant level of CO_2 observed in the regenerator gas during this period.

It is evident from regenerator performance during this run that regenerator efficiently is strongly affected by sulphur concentration in the incoming lime and that high sulphur concentrations improve regenerator efficiency and reduce the relative amount of sulphur recycle.

5. End of Run Inspection

At 1700 hours on October 11, gasification was terminated and the bed was converted to sulphate with a continuous flow of air and recycle gas (while adding stone to control temperature).

The unit was held at temperature by gas firing for approximately 24 hours and then was shut down and allowed to cool.

When opened, the gasifier was found to contain no lumps or solid deposits such as those observed around the fuel injectors after the first test. Gas transfer ducts, cyclones and the lime circulation passages also were clean. Of course, the period of hot operation after terminating gasification would have removed any carbonaceous deposits but the absence of lime agglomerates in the gasifier and transfer lines shows that the use of nitrogen rather than air for pulse gas in the transfer system was effective in avoiding agglomerates.

The lower portion of the regenerator did contain a single large deposit which covered the centre portion and about half the perimeter of the distributor. This deposit tapered upward and away from the port through which solids entered from the gasifier. Inspections showed that the distributor plate was not well seated at its base, and evidently had bypassed a large portion of the fluidising gas around one side of its rim. It appears that this bypassing had reduced the pressure drop through the distributor holes enough to permit defluidisation on one side of the distributor. The defluidised lime particles then had agglomerated, forming the deposit found at the end of the run.

It is evident that incomplete seating was achieved when the distributor was replaced under hot conditions just before starting gasification. Care in seating the distributor together with provision of an air flow to the regenerator throughout heat up should avoid this problem in the future.

A considerable quantity of fine lime was found in the flue gas recycle line, blower, and in the gasifier plenum. However, no obvious pressure drop increase had occurred in these systems during the course of the run.

C. UNIT MODIFICATIONS PRIOR TO RUN 3

A fairly extensive series of modifications was made to the pilot plant prior to Run 3 to correct problems which had arisen.

1. Gasifier

a. Bed Feed System

A modification was made so that the limestone addition would be introduced directly into the fluidised bed through the sidewall rather than dropped through the lid with the possibility of some of the finer material being blown straight out before reaching the bed. Due to limited time this system could not be developed to give complete reliability and so the original system of gravity feed through the lid was left available in case of trouble.

b. Fuel Injectors

The hole in the end of the fuel injectors was increased from 3/16" dia. to 1/4" dia. to reduce the fuel and air velocity.

c. Cyclone Tubes

Double walled type 310 stainless steel cyclone tubes were fitted with provision for steam cooling. Each tube was instrumented in the interspace with thermocouples.

2. Gasifier Services

a. Flue Gas for Decoking

An additional piping system was added so that the flue gas could be directly recycled into the gasifier top space through the lid to provide a means of controlling the carbon burn-out in the ducts.

b. Flue Gas Recycle Cyclone

A cyclone was fitted on the flue gas recycle line to reduce the dust returned to the blowers.

c. Outlet Gas Clean-up

A knock out chamber followed by a cyclone was built into the main flue between the boiler and stack to reduce the stack solids emission.

3. Test Specimens

Arrangements were made to obtain and test two alternative materials under gasifying and controlled burn-out conditions with the aim of determining their suitability for possible use as cyclone outlet tubes. Silicon nitride, and self bonded silicon carbide test pieces were fixed in the bifurcated duct above the outlet from each cyclone.

D. CAFB - OPERATIONS LOG RUN 3

19.11.71 The propane gas burner was started up at 18.00 for the gradual heating-up process. Some hot gas leakage around the lid but these were sealed with fireclay.

20.11.71 The heating continued on schedule.

21.11.71 A test was made of the regenerator destributor flow/pressure drop characteristics.

The gas burner failed and the flame detector was cleaned and replaced.

A blocked pressure tapping in the regenerator was cleared.

22.11.71 General checks were made on pressure measurement and the bed pulser control equipment. The regenerator to gasifier pulse controller gave only a continuous signal instead of an intermittent one and a wiring fault was detected and rectified.

The three metering pumps were started on kerosene. The No.2 pump kept cutting out and affecting No.1 pump which was on the same electrical circuit. No.2 pump was isolated and the unit ran on pumps 1 and 3.

Lime feed was started but the modified direct injection feed system was inclined to block. The original gravity feed through the lid system was replaced.

23.11.71 The limestone feed was continued, but there was some difficulty in maintaining the bed temperature.

The Compton compressor broke its connecting rod bolt and was stripped down to await new components.

The unit was shut down for a short period to permit a modification to the electrical supply for the metering pumps to give them individually fused supplies.

The first stage blower drive pulley became loose and threw a belt. Pulley was relocated and the blower restarted.

24.11.71 There was still little progress on getting the bed to fluidise. The kerosene was burning above the bed and modifications to the fuel injectors did not help in the combustion within the bed.

It was considered that the bed material may have had insufficient fine particles for fluidisation at available gas velocities. Some lime was reground and gradually fed into the unit after some of the bed had been drained off.

25.11.71 Gradually the bed began to fluidise, and reground limestone was successfully added to build up bed height. The regenerator air rate was increased but fluidisation and transfer was not good.

The nitrogen system for bed transfer and general inerting requirements was checked out, and it was discovered that the suppliers had omitted their daily visit to refill the bulk nitrogen storage which by now was almost empty. The gasifier was switched over to air for bleeds and pulsers to conserve nitrogen for other inerting requirements on the site when nitrogen was essential.

The Compton compressor was repaired and checked out satisfactorily.

26.11.71 The left hand cyclone had been deliberately plugged at its lower end where it drains to the regenerator to gasifier transfer duct, to prevent back flow of material during the bed filling operation. Now that the bed height was up to its operating level the plug was withdrawn so that the cyclone fines would be returned to the gasifier bed.
After various adjustments to pulser pressure and bleeds, the transfer system operated well together with good fluidisation in the regenerator.
At 05.30 the liquid nitrogen storage tank blew its safety valve which failed to reseat and caused loss of the tank contents. The liquid N_2 suppliers came and rectified the valves, repiped the system to ensure that the bursting disc would operate independently of the safety valve, and refilled the system. The cause of the trouble was shown to be an incorrect filling procedure by the tanker used the previous evening. The boiler extension tube was removed and the boiler rear door closed. Some trials were made on the main flame pilot. By starting up at low air flow rates to the burner the pilot was established and could be maintained while the air rate was gradually increased.
All external cyclone dust collection drain systems were installed and sealed to minimise dust emission during collection.
The cyclone cooling steam lines were completed and checked out.
At 19.30 a low cooling water pressure switch shut the plant down. Examination showed that the soft water make-up flow meter had seized up and cut off the flow of cooling water. The meter was by-passed and the supply resumed.
The burner-to-boiler joint was sealed with fireclay and asbestos rope after the unit and ductwork had reached their operating conditions.

27.11.71 (Day 1) The analytical sampling system was completed in readiness for gasification.
At 01.30 gasification started - the main flame ignited first time - as usual.
The first few hours were spent in getting the unit to settle down and the transfer rate of bed from the gasifier to regenerator was regulated to obtain good regeneration.
A boiler fire tube temperature of 1086°C was measured by suction pyrometer through the boiler rear door.
After 22 hours of gasification the regenerator SO_2 level had reached 9.6% . The automatic regenerator temperature controller was cut in which held the temperature steadily within 5°C of the set point.

28.11.71 Day 2) A further measurement of the boiler fire tube back end temperature showed an increase to 1139°C over the first 24 hours of operation. The gasifier bed had gradually been increased to 19" and associated with this then was some difficulty in controlling the regenerator temperature - possibly due to a higher rate of bed transfer. The nitrogen pressure in the pulser system was reduced and this helped to line out conditions.
The rate of transfer was very sensitive indeed to the relative pressure between gasifier and regenerator. Irregularities in gas space pressures produced irregularities in bed transfer which in turn upset the regenerator temperatures.
The regenerator drain would not run slowly at the required rate of 15 lbs/hour and intermittent manual removal was necessary. The

drain was rodded out before each removal to clear away a crust in the entry to the drain.
The silicon nitride test specimen showed some irregular temperature variations giving values up to 1090°C.
The gas sampling flame was tried out but there was not sufficient gas pressure to give a good flame. A Drager tube test on the boiler flue gas gave a value of 100 ppm for NO_x.

29.11.71 (Day 3) Some plugging was experienced with the limestone bed feed system and very delicate balancing of the pressures between the gasifier and limestone hopper was necessary to ensure a steady flow.
A manometer was added to the boiler flue to provide information on the pressure drop across the boiler water tubes.
Gradually the gasifier gas space pressure had risen to 17" and with the blower fully open any subsequent pressure increases cut back on total air and flue gas to the gasifier.
The gasifier distributor has shown a gradual increase in pressure drop since the start of the run.
At 18.30 the bed was sulphated by controlled oxidation using full recycle of flue gas and limestone feed to control the temperature.
At the completion of sulphation the bed was slumped and a controlled carbon burn-out was completed using flue gas recycle and nitrogen injection into the gasifier top space. The cyclone tubes were cooled with steam to prevent them overheating. The oxygen concentration was controlled by air addition to the boiler through the burner secondary air supply.
The carbon burn-out was completed and gasification resumed at 22.05.

30.11.71 (Day 4) The next few hours were spent in letting the unit settle out but some irregularity in the gasifier and regenerator pressures produced fluctuations in the bed transfer which in turn caused variations in the regenerator temperature.
It had become apparent that the limestone feed had a considerable quantity of material larger than 1/8". At this stage it was decided to pass the material through an additional sieve to remove these larger particles to prevent their accumulation in the lower part of the gasifier bed.
The gas space pressures had risen to 16.5" and preparations were made to start another carbon burn-out which commenced with sulphation of the bed at 21.18 followed by a burn-out. The silicon carbide test specimen rose in excess of 1200°C whilst the silicon nitride specimen temperature became erratic.

1.12.71 (Day 5) At 01.20 gasification was resumed and conditions soon steadied out.
Trials on varying the air rate to the cyclone bleeds produced downstream temperature changes particularly on the silicon carbide specimen. The thermocouple in the silicon nitride sample stopped functioning reliably.
Conditions steadied out with an optimum pressure differential of about 5" between gasifier and regenerator which gave a steady controllable transfer rate.
The limestone feed rate was high at this period showing about 70 lbs/hr with only a small increase in bed levels. The corresponding fines removal rate from the regenerator cyclone was also higher than normal and possibly this batch of limestone was dustier than usual.
At 08.20 some difficulty was found in draining the regenerator - the skin across the drain seemed very hard and had to be hammered to break it away.

The bed feed became sluggish. Replacing the feed vibrator controller improved matters.

At 16.30 there was some difficulty with the regenerator temperature controller. When temperature exceeded the set point by too great a margin, pulsing stopped and a steady injector gas flow resulted. This action caused a decrease in solids circulation rather than the desired increase. The remedy was to increase temperature set point momentarily to restore pulsing and to increase gasifier to regenerator pressure slightly which increased maximum solids flow rate.

Conditions were steady with excellent sulphur removal in the gasifier and good regeneration but gradually the gasifier pressure had risen up to 20.5" with corresponding decrease in the air and flue gas flow rates.

Preparations were made for a further burn-out by the usual technique of limestone addition and flue gas recycle for sulphation followed by a controlled carbon burn with the gasifier bed slumped.

The burn-out started at 17.34 and at 19.16 the unit was shutdown to permit draining of the left hand cyclone prior to removal of the silicon carbide test specimen located above this cyclone outlet. It was thought that this hanging specimen could be acting as a collector for carbon and lime and when removed some of this material might fall into the cyclone and block the fines return line. Gasification was resumed at 21.30.

2.12.71 (Day 6) Unit steady after the burn-out apart from some temperature variations in the gasifier (a). These temperature swings were accompanied by 30°F variations in the fuel oil preheater temperature. The burner throat temperature jumped up to 1200°C for a few hours. The bed feed vibrator again became sluggish, and it was replaced by one removed from the regenerator drain. This did not help much in improving the rate of bed feed.

At 06.30 the gasifier pressure had again built up to 19" w.g. with the blower fully open, and another burn-cut was necessary. Due to unreliable vibrator, this burn-out was done without the cooling aid of limestone feed into the gasifier. Previously the regenerator air flow had been reduced after the sulphation of the gasifier bed to about 3 cfm during the burn-out period. On this burn-out, which started at 06.55, it was decided to leave the regenerator fluidised but this caused a difficulty in that the air supply came up the transfer duct into the gasifier and upset the control of the gasifier during the carbon burning. The air rate was reduced to 9 cfm and so re-established the differential pressure between gasifier and regenerator to stop the back flow of air.

In spite of the absence of cooling limestone addition this burn was controlled with flue gas recycle, nitrogen addition and steam to the cyclone tubes.

During the burn many red hot particles and sparks were seen passing into the boiler.

These last two burn-outs were short in duration with small pressure recovery which indicated that the amount of carbon was probably small and the pressure rises were most likely due to inorganic material.

(a) Temperature variations later traced to uneven fuel rate caused by low tank level.

A sieve analysis of the feed limestone showed it to have a major portion in the 1180 to 1400 micron range with a very small quantity greater than 1400 microns. There were considerable variations in the limestone from batch to batch.

Whilst still on combusting conditions before resuming gasification, the first stage blower was run to provide an additional boost of air with the intent of blowing some of the lime deposits out of the system.

At 11.05 gasification was resumed, but the gasifier distributor pressure drop was 23" leaving very little margin for additional resistance from further duct obstruction. Plans had been put in hand earlier to rerun some of the pipework so that both blowers could be used with flue gas recycle, and at 15.20 the unit was shut down to permit final connections to be made.

During this work the bifurcated duct was rodded out from the rear access plates on each side and some material fell down into the cyclones.

At 18.33 the unit was put back on gasification with both blowers operating. It was observed that there could be some relationship between pressure rise on the ducts and high gasifier temperature and, therefore, the gasifier temperature was lowered to 870 - 880°C. The cyclone air bleed rates were cut back to a lower value because of the possibility of the higher temperatures induced by these bleeds causing the fines to pass through a sticky phase and so accelerate the rate of pressure rise.

The bed depth was built up slowly but the vibrator was not completely satisfactory, and this impeded the rate of bed addition.

3.12.71 (Day 7) There was a gradual rise in the gasifier pressure of 1" in 3 hours but otherwise all ran quite steadily with the new blower system.

At 11.20 some samples were taken from the top of the stack for particulate emission tests and NO_x samples taken from the boiler.

About mid-day the SO_2 level began to rise in the boiler. There had been a slow bed feed into the gasifier in the preceding hours, and so some material was withdrawn from the regenerator in order that fresh bed could be added without increasing the bed depth. After about 2 hours, the SO_2 level had dropped back in the boiler.

During the day two further particulate emission samples were taken from the top of the stack.

The rate of pressure rise was the greatest across the cyclones and a further short burst of air was added through cyclone probes. There was some very slight improvement in the pressure drop in this area.

The limestone feed again proved erratic and the surface bed level fell during the evening with corresponding increases in the boiler SO_2 concentration.

The regenerator cyclone had not collected much solids since the previous burn-out and it appeared that the gasifier cyclones may have become blocked when the bifurcated duct was rodded in the morning.

4.12.71 (Day 8) The gasifier distributor pressure drop continued to rise and the air bleeds with the fuel were increased to break up any larger bed particles which could have segregated to the bottom of the bed and contributed towards the higher pressure drop. There was not any significant improvement with this change in fuel air rate.

The increased quantity of material removed from the boiler and its cyclone suggested that the gasifier cyclones were not effective at that time.

Steam injection to the cyclone inlet ducts was tried for a brief

period to determine whether this technique was more effective than air injection but there was no significant change.
The gasifier to regenerator solenoid valve failed, and the standby valve was switched into circuit.
At 14.55 the bed was sulphated prior to another burn-out. The burning had not finished after 82 minutes, but because of the drop in the bed temperature the unit was put onto combusting conditions. Both the cyclone drains were emptied out, giving 63 lbs. of fines from the right hand cyclone and 40 lbs. of fines with some lumpy material from the left hand side.

5.12.71 (Day 9) The boiler SO_2 level rose during the period that the limestone feed was shut-off for hopper refilling.
The regenerator temperature control became erratic for a few hours apparently caused by pressure fluctuations, which in turn caused irregular bed transfer.
The boiler tube pressure drop continued to rise. Attempts were made to drain the various outlets of material but without much success.
The regenerator drain was found to be blocked at 06.45 and all attempts to clear it were unsuccessful apart from allowing small quantities of material through. Some material was withdrawn from the gasifier in order to maintain the fresh bed feed.
During the course of the morning it was decided to try and get the right hand cyclone fines return line functioning so that the dust emission to the boiler could be reduced. The unit was sulphated at 11.15.When this was complete, the bed was slumped whilst the cyclone was drained and the gasifier to regenerator transfer line drained off to provide access for rodding out the fines return line.
The silicon nitride test specimen holder situated in the bifurated duct over the right hand cyclone was removed, and it was possible to see into the red hot cyclone and observe the rod come up the fines transfer line.
At 15.28 gasification was resumed but still there was not much dust collected in the regenerator cyclone.
The back end of the boiler which had previously been bypassing some hot gas due to the pressure rise across the fire tubes seemed to improve during the cooldown and behaved reasonably after this period.
The gasifier bed level gradually rose to 17" and boiler SO_2 level remained below 100 ppm although regenerator SO_2 level was slow to rise after the burn-out.

6.12.71 (Day 10) The bed feed rate was set to just maintain a steady bed level and with this slow feed the rate of rise of boiler tube pressure was steadied.
The oil fuel tank was switched to the full compartment, and the additional head produced an increase in oil flow rate which in turn caused the burner in the boiler to go sub-stoichiometric for a few minutes. Under these reducing conditions the sulphate in the boiler liberated SO_2.
The unit continued to run smoothly and with a bed feed rate around 1.5 wt. CaO/wt of sulphur the removal efficiency was greater than 90%.
As the boiler fire tube pressure drop continued to rise very slowly the air and fuel rates were backed off to compensate.
At 16.30 the unit was shut down at the completion of over 200 hours gasification.

E. Examination of the Continuous Gasifier and Auxiliary Equipment after Run 3.

At the end of the run, at 1630 on Monday, 6th December, the unit was shut down by stopping the fuel pumps, followed by all blowers as soon as the main flame had gone out. Not enough air was put into the unit at this time to sulphate the bed or burn any carbon from the system. The unit was then sealed and allowed to cool. The intent was to leave the unit with any deposits in place as they had been at the end of operations. It is apparent, however, from the trace of the temperature recorder which was left operating, that about $3\frac{1}{2}$ hours after shutdown some air entered the gasifier and burned off much of the carbon residue. The gasifier gas space thermocouple which had fallen to 500°C, began rising and reached 570°C. The bed temperature also stopped falling for a period of about 10 hours as heat was evolved. It is probable that the air entered around the gasifier lid and leaks developed during cooling and that a draught was set up through the boiler and regenerator outlet flues.

1. Gasifier

a. Gasifier Distributor

The nozzles of the distributor had shown an increasing pressure drop from 9.5" water gauge to 33.8" during the duration of the test. Examination showed that all the nozzles had some degree of obstruction both within the bore and sometimes at the exit as well. 20% of the holes had at least 25% of their area obstructed and about 2% were completely obstructed. The obstruction in the bore was formed by fine material adhering around the periphery of the bore, and the external obstruction was caused by a thin flaky deposit attached to the stainless steel nozzle. A photograph of several nozzles appears in Figure J-36.

The fine material deposited in the holes of the nozzles was carried in the flue gas recycle and thrown out of the gas stream by changes in gas direction and turbulence at the nozzles. The flaky material round the outside of the nozzles could have been formed from fines within the bed being locally regenerated by the air from the nozzles and so passing through a sticky phase at the higher temperature and adhering to the nozzle.

The distributor refractory covering was coated mainly around the periphery with a whitish deposit which in some areas was thick enough to obstruct the holes in the distributor. In other areas the air from the nozzles had cut a channel in the deposit which may be seen in Figure J-37.

Apart from the deposits mentioned above, the refractory concrete was in good condition with one or two hairline cracks and a few chipped edges.

b. Gasifier Refractory Concrete

(1) Surface Condition

Generally the surface finish had not deteriorated very much since the first room temperature fluidisation trials in which the surface became quite rough particularly immediately above the bed. The rough surface is not uniform around the gasifier faces and could be caused by the method of pouring and vibrating the concrete which caused a concentration of large particles near the surface which were then easily removed by abrasion.

During Run 3 a black flaky deposit was laid down on the lower gasifier faces, particularly on the short chamfered faces at the corners. Fig J-37 shows a light coloured zone on the bottom of the gasifier walls which dips away around nozzles which were blocked and at a regenerator to gasifier materials return line. This colour change could indicate the extent of the oxidising zone present in the gasifier with the dips corresponding to regions without oxygen.

The refractory concrete was cracked down both the gasifier short sides after the critical firing process, and these two cracks have increased in width a little during subsequent operations. The worst crack is in the lowest concrete lift which is around the cold penum. Here the concrete is put into tension by the expansion of the hot concrete immediately above it. The cracks in the hot zone will almost close up at operating temperature but those in the plenum will stay partially open. The gasifier distributor and plenum is one integral subassembly and hence there cannot be any leakage of air through this crack to bypass the distributor. Because of air leakage during cool down, walls of the gasifier gas space were relatively clean.

The lower area of the gasifier was blackened to some extent and the two cracks in the walls had a carbon and lime deposit around them.

There were some other small cracks on the faces which often terminated at the junctions between the various lifts in the concrete. This limitation of cracking shows the usefulness of this form of construction.

(2) Start-up Burner Quarl

Flame from the start-up burner enters the gasifier through a downward angled, precast, refractory tube or quarl.

There was a large deposit of lime particles in this quarl, most of which fell out once the lower portion was moved. Some finer material had adhered to the walls especially adjacent to the burner itself.

The lower section of the quarl developed a crack during the last run but the piece was still firmly held in place by the concrete around it.

(3) Gasifier Lid

The concrete slabs on the gasifier lid were generally clean apart from a build up of carbon and lime around the limestone feed point. A few fragments of concrete had broken away from the hot face of the concrete but as these were lying on top of the slumped bed it evidently happened during cool down or whilst the lid was being lifted out. There was a considerable quantity of carbon and lime deposited in the annulus between the lid blocks and the gasifier wall.

(4) Materials Transfer Ducts

- Gasifier to Regenerator

This duct remained in good condition with no signs of any blockages. The refractory became a little brown during the course of all the operating time but

- Regenerator to Gasifier

It was not possible to inspect all the internal bore of this transfer line but there was a small deposit of agglomerated fines on the flat rear face of the transfer line. This deposit was covered by a skin which presumably formed when the material passed through a sticky phase. The composition of a sample of the deposit is listed in Table J-IV.

The entry to this transfer line in the regenerator wall was partially obstructed by a build up of fine white material on the regenerator wall which spanned across some of the opening.

The refractory concrete was generally good with no evidence of cracking or spalling.

c. Gasifier Bed

Figure J-38 shows the gasifier bed after removal of the lid. The debris on the surface fell from the lid or the lid joint either during cool-down or actual removal of the lid. Most of the carbon particles were trapped between the lid concrete blocks and the gasifier wall.

The bed was uniformly withdrawn by vacuum tube until three dark zones could be seen corresponding to the fuel injector outlets. The bed surface at this level appears in Figure J-39. These darker areas were probably formed at shut down by oil cracking on the hot slumped bed - the inequality in the size of the zones could possibly be caused by the position of each metering pump in its stroke at shutdown.

There was not any visual evidence of particle size segregation in the bed but sieve analysis of the lower portion of the bed showed 35.6% greater than 1400 microns whilst only 26.6% of the upper bed portion was greater than 1400 microns.

2. Regenerator

a. Distributor

Figure J-40 is a photograph of the regenerator distributor after the run.

The distributor was constructed from refractory concrete with sixteen $\frac{1}{4}$" diameter holes. After the 200 hour run when the distributor was removed a piece broke off at the corner. Possibly a crack formed during the run, but when the distributor was in place it was held tightly together and would not have bypassed any significant quantity of air. Six of the

Table J-IV

Composition of Solid Samples Removed from Pilot Plant After Run 3

Sample Location Composition		Deposit From Regenerator to Gasifier Transfer Line	Deposit from Bottom of Regenerator	Annular Deposit from Regenerator	Deposit from end of No.2 Fuel Injector	Bed by No.3 Fuel Injector	Obstruction from L.H. Cyclone Solids Drain	Inner Core from Fire Tube Deposit	Outer Surface of Fire Tube Deposit	Flue Gas Cyclone Drained Fines 198½ Hr.	Regenerator Cyclone Fines 198½ Hr.	Flue Gas Cyclone Fines 198½ Hr.	Boiler Drain Solids 198½ Hr.
Loss on Heating to 900°C,	wt %	(2.39)(a)	1.78	0.77	-	-	19.93	1.96	2.34	6.58	3.18	4.55	3.41
Total Sulphur	wt %	22.84	17.95	20.09	-	-	17.94	20.38	20.20	4.01	9.65	3.79	5.86
Sulphate Sulphur	wt %	13.44	16.20	19.16	-	-	-	20.38	20.20	3.07	7.06	3.01	1.93
Vanadium	wt %	0.35	0.67	0.17	-	-	-	0.32	0.27	0.60	0.25	0.63	0.83
Carbon	wt %	-	-	-	67.2	4.15	54.06	-	-	-	-	-	-
CaO	wt %	-	-	-	-	-	-	-	-	-	-	-	-

(a) wt. gain

holes were plugged with a fine deposit of white powder on the inlet side and loosely plugged on the outlet end with coarser material. Examination of the records shows that the pressure drop in the distributor was about $1\frac{1}{2}$" before any bed was added to the gasifier but rose to 6" when the bed was added suggesting that some obstructions were formed during the bed addition. During this period the air flow to the regenerator was kept below 2 cfm to reduce the cooling effect upon the regenerator refractory. This air rate may not have been adequate to prevent blockage of some of the holes.

The top face and edges of the distributor showed signs of abrasion with the removal of the finer material leaving the larger particles exposed.

b. Regenerator Bore

The bore was partially obstructed by a block of agglomerated fines with local pockets of coarser material inbedded in it. The obstruction was strong and showed some layering in its construction as if it had built up, formed a surface skin, and subsequently had another layer deposited. Figure J-41 shows a view of this deposit. The analysis, listed in Table J-IV, indicates that it is largely calcium sulphate. It contains enough vanadium to suggest that it contains solids from late in the run and was not formed just at the run beginning.

Higher in the regenerator there was a further whitish annular deposit shown in Figure J-42. It was about $\frac{1}{2}$ inch thick and extended about 14 inches in height. It partially obstructed the entry to the solids transfer port from regenerator to gasifier. Its composition (Table J-IV) also corresponds with calcium sulphate. Its lower vanadium content suggests that it was formed from fines which had not remained in the gasifier long enough to pick up much vanadium. Above this annular deposit there were local areas of white deposits in the regenerator and in the refractory lined ducts leading from the regenerator. Figure J-43 shows these deposits at the regenerator outlet.

In previous tests there had not been any evidence of deposits in the outlet ducts. These deposits were most probably held together by some of the fines in the bed material passing through the sticky phase and acting as a cement to bind other fines and coarser material.

c. Regenerator Concrete

There were some small cracks in the concrete walls but it is unlikely that any leakage could have occurred at running conditions. The cold zone around the plenum had two cracks which had been filled with fire clay prior to the test run, and these have remained well sealed during the test.

3. General Penetrations

The penetrations into the gasifier and regenerator fluid beds for temperature and pressure measurement were formed of 27% chromium type 446 steel which withstood the conditions with some scaling. These components were not replaced during the total of approximately 460 hours of gasification and many hours of combustion.

The pulsers for bed transfer control also fabricated from the same material were in excellent condition but have been cooled by the nitrogen passing through them.

The stainless steel fuel injectors and bed feed tubes were undamaged. The former were cooled at all times by either fuel or nitrogen, and the latter were set back into the refractory concrete to minimise heat pick up.

The fuel injectors had black shiny deposits around their exposed ends, and the centre injector had a 2" long extended shroud of black material which had built up at its end. This deposit contained 67% carbon. The two air injection tubes which passed through the lid and extended in front of the cyclone inlets were used to blow air into the inlet to raise the temperature in an effort to reduce carbon deposition. This technique certainly affected downstream temperatures, but there was no clear evidence of it being effective in reducing carbon deposition. One trial was made with steam injection in one tube. Subsequent examination showed that the tube end had broken away and dropped into the gasifier bed. These tubes were both heavily scaled.

4. Gasifier Cyclones

The right hand cyclone was rodded out towards the end of the test run, and apart from a small deposit near the bottom which was not removed, the cyclone and its fines return line were quite clear. (Figure J-44)

The left hand cyclone had not been cleared out and was blocked to a level of about 13" above its drain off point (Figure J-45)

The material in the cyclone consisted mostly of fine black free flowing particles in the upper layer with a few larger pieces of agglomerated fine lime pieces and carbon lying on this layer. As the material was removed the colour gradually lightened until at the bottom of the cyclone it was quite white.

This accumulated material was prevented from flowing into the open transfer line by a very hard piece of agglomerated white fine material about $2\frac{1}{2}$" x 2" x $1\frac{1}{4}$" which was lodged across the entry of the fines return tube. It is not clear how this obstruction formed but it is unlikely to have fallen from the bifurcated duct because experience shows duct deposits to be less dense and hard. The analysis listed in Table J-IV corresponds with a composition of 76.5% Ca SO_4, 18.9% CaO and 4.5% CaS.

The cyclone entries, shown in Figure J-46, unfortunately had been burned reasonably clean leaving some granular deposits at the bottom of each duct. The entires to these ducts were radialized after the first run and made smooth by buttering with a fine refractory paste which stayed in place.

The cyclone gas outlet tubes were redesigned following the melting of the tubes in the first run. Double walled stainless steel (type 310) tubes were used. Each tube was divided into 2 segments through which steam could be passed. An internal guide was incorporated to direct the steam down one segment and return up the other. The outlet steam vented to atmosphere.

The tubes were tested to 5 psi prior to installation and after removal retested. The left hand tube was scaled quite heavily but still was leaktight. The right hand tube was also scaled but leaked around the lower welded joint between inner and outler skin. Close examination showed a line crack in the weld zone. There is some differential expansion between the inner and outer skins due to the outer skin being half embedded in refractory concrete and not being in contact with the hot gas. This expansion would give rise to bending and tensile stresses in the joints between inner and outer skins which together with some corrosion could be the cause of this failure.

The double walled tubes were both instrumented internally with four chromel-alumel thermocouples located at various levels within the tubes. In the left hand tube 2 of the thermocouples stopped working during the test but the right hand tube instruments were satisfactory all the time.

5. Bifurcated Duct

The hot gas ducts leading from the two cyclone outlets became partially obstructed at the junction between the vertical outlet from the cyclone and the horizontal duct to the burner. This junction included a right angled change in direction where material was thrown out by the gas to accumulate on the corner. Examinations were made by removing the end access covers, and at one of the burn out operations the ducts were rodded clear. It was during this operation that material fell into the cyclones. Another contributory cause of deposits in this area was the test specimens of alternative cyclone outlet tube material which were hung in the bifurcated duct at this junction. These specimens acted as collectors for material and therefore one specimen was removed half way through the run whilst the other one was accidently broken during the rodding operation.

There was some accumulation of carbon/lime deposits at the junction between the two horizontal ducts prior to their entry into the main burner section.

The expansion joints between the gasifier and the bifurcated duct were in good condition without any evidence of cracking or deposition of material.

6. Burner

a. Premix Section

The premix section was constructed from type (310) stainless steel in the hot zone with ceramic fibre insulation. There was some flaky 1/16" thick carbon deposits on the bore of the inner hot duct but no evidence of any deterioration.

b. Main burner

This burner was constructed from stainless steel in the hot zone surrounded by an external annulus of cooling air from the secondary air supply. The thermocouple which was inserted into the hot gas stream gave erratic readings after 160 hours of gasification, and the protruding end of the stainless steel clad thermocouple had completely disappeared at 200 hours. There were times during the operation, particularly during and after a burn out when this temperature apparently went above 1200°C for some hours probably due to local carbon burning off. During gasification there is some oxygen available in this zone from the premix air supply.

There was a thin deposit of carbon/lime in the burner hot gas duct but no evidence of any build up of material in the burner throat.

7. Boiler

a. Fire Tubes

The front end refractory around the burner has not deteriorated during the test, and the stainless steel deflector plates in the pilot flame area were in good condition. There was a deposit of fine lime around the area of the boiler together with deposits in the boiler corrugated fire tube.

At the rear end of the fire tube there were very local white deposits (Figure J-47) adhering to the gas tube in random positions. It is assumed that these were formed from fine material in the flame which went through the sticky phase and gradually built up at random locations. The deposit was hard but could be removed fairly easily. The greatest deposits were at the entrance to the first tube pass where the majority of the tubes were partially obstructed by a hard thick crusty coating of fines. In some cases the tubes were completely obstructed. The crust could be broken off fairly easily and left a cone shaped deposit in many of the tubes which could be easily withdrawn. The longest plug removed was 15" with the majority less than 6" long. Figure J-48 shows the largest of these cores. The core composition (Table J-IV) corresponded with a high content of calcium sulphate. The vanadium content is in line with that expected for the average lime to vanadium feed ratio.

The gas exit ends of the tubes of the final pass did not show any crusty formation but many of them, particularly in the lower portion had a deposit of fine material which was easily removed.

Figure J-49 is a general view of the open boiler backend. The firebrick arch separating the passes in the boiler showed some flaky deposit of lime, and on the horizontal section beneath the main outlet there was a deposit of finer material. The asbestos seal between boiler passes looked reasonable although there was evidence of gas bypassing near the end of the test run due to the build up of pressure across the tube passes. The boiler paint work was discoloured at the lower portion due to this hot gas leaking through the internal seal and overheating the casing locally.

b. Stack and flue gas cyclones

The flue gas cyclone was added to the system before the third run to reduce particulate emissions and the system worked well. The stack was clear with a very small deposit of lime particles on the horizontal section at the bottom of the stack.

The flue gas recycle cyclone was also added to reduce the fines returned to the blower system. During the run material was drawn off but some fine material was still passing and causing distributor blockages in the gasifier.

8. Test Specimens

After the experience in the first gasification run when the stainless steel cyclone outlet tubes were melted, investigation was made into the use of alternative materials. Two refractory materials were chosen for sample tests by hanging the specimens in the bifurcated duct just above the cyclone outlet with thermocouples placed under the specimens.

In this run the temperature during burn out was controlled by using flue gas recycle together with nitrogen injection to supplement the flue gas if required. The oxygen level was controlled by passing air into the system at the burner. This air returned to the top of the gasifier with the recycled flue gas.

The sample temperatures could not be controlled other than by the gas recycle whilst the stainless steel cyclone tubes were steam cooled as well.

a. Silicon Nitride

A closed ended sheath of silicon nitride was fixed above the right cyclone outlet. The recorded temperatures in this sample were rather erratic. Initially the temperature lagged on the main gas stream but then exceeded the gasifier top space temperature. Much of this was caused by the air injection into the cyclone inlets and subsequently erratic behaviour arose during and after carbon burn outs when the indicated temperature rose above 1200°C.

When the bifurcated duct was rodded clear at the second burn out on day 6, the test specimen was broken and fell into the cyclone. No trace of fragments was recovered. A similar specimen which was tested in the second run became very brittle. The results from an analysis by the manufacturers show that the lime reacted to form a mixed metal silicate which leached out the silicon leaving a brittle material.

b. Self Bonded Silicon Carbide

Self bonded silicon carbide is a recently developed dense material available from the U.K. Atomic Energy Authority. Our test specimen was a $\frac{3}{4}$ inch o.d. x $\frac{1}{2}$ inch bore open ended tube. This tube was hung above the left hand cyclone in the bifurcated duct with a thermocouple fixed inside. The specimen temperature again showed considerable variations with the gasifier top space temperature - particularly after a carbon burn out operation.

After 104 hours of gasification in Run 3 the specimen was removed intact to prevent its accumulating lime and carbon and contributing to duct pressure drop. The tube appeared by visual inspection to be undamaged. It was returned to the suppliers for more detailed examination. Some strength tests were conducted on test sections cut from each end of the specimen, and these were compared with a control specimen. Sections were also cut and examined under the microscope again comparing with a control specimen. The manufacturers say that the material is unaffected by its exposure to gasifier conditions.

9. General Service

a. Boiler

The boiler and its associated cooling system have been trouble free while maintaining a steady boiler water temperature throughout the test.

b. Oil Supply

The rate of oil supply to the gasifier became a little irregular when the oil quantity in the external oil storage tank became lower than 500 gallons. The system acted as though the circulating ring main began picking up air bubbles at this tank level.

c. Gasifier controls

The control, analytical and recording equipment functioned reliably apart from some minor electrical troubles.

d. Bed Handling

The alternative method of direct injection of the bed material was not sufficiently reliable due to the limited time available for its development. The use of a vibrator for controlling rate of bed feed was unreliable for consistent input because it was very sensitive to the end constraints of the pipework. Investigations are in hand for alternative methods of bed feed control and input.

The vibratory regenerator drain did not function as a steady slow method of bed removal because the pipe became plugged unless a fast flow of material was maintained all the time. Investigations are in hand to improve the present system which in this run relied upon manual draining at intervals.

e. Blowers and Compressors

The gradual rise in the system resistance entailed modifying the blower pipework so that the flue gas recycle system could be used with both blowers in series.

f. Limestone

It is not practicable to prepare locally the quantities of limestone required with the smallest particle size below 800 microns due to sieve blinding caused by the dampness of the material as received from the quarry.

F. Computer Programmes for Pilot Plant Data

Several programmes were written in BASIC language for the G 265 time sharing computer system to aid in processing and analysing CAFB pilot plant data. These programmes are summarised and listed here for reference.

1. Data Filing Programme

Programme JHTPPF was written to put a number of pilot plant data items into file for future reference. It was intended that the process operator would update the files periodically with hourly sets of data readings. Because of file size limitations in the G-265 Mark I system it was necessary to string several files together and to build into the filing programme instructions to keep track of which file was in use and the location of the last data item filed.

The data are entered with the names D(I) where I goes from 1 to 25. The individual data items have the significance listed in Table J-V below.

Table J-V

Data Listed by Pilot Plant File Programme

Data Designation	Data Item
D (1)	Day. Hour
D (2)	Gasifier Temperature, °C (Top of Bed)
D (3)	Regenerator Temperature, °C (Top of Bed)
D (4)	Gasifier Air to Blower, SCFM
D (5)	Flue Gas Recycle, SCFM
D (6)	Air to Injectors and Start Up Burner, SCFM
D (7)	Air to Stone Feed, SCFM
D (8)	Pressure in Gasifier Above Bed, in.w.g.
D (9)	Pressure at Regenerator Air Meter, in. w.g.
D (10)	Pressure at Regenerator Nitrogen Meter, in.w.g.
D (11)	Regenerator Air Flow, Meter CFH
D (12)	Propane to Boiler Pilot, SCFM
D (13)	Oil Rate, Gallons (Imp)/Hr.
D (14)	Oil Meter Temperature, °F
D (15)	Nitrogen to Regenerator, (Meter CFH)
D (16)	Stone Feed Rate, lb/hr.
D (17)	Oxygen in Plenum Gas, Vol %.
D (18)	Oxygen in Boiler Flue Gas, Vol %
D (19)	Oxygen in Regenerator Gas, Vol %
D (20)	SO_2 in Boiler Flue Gas, ppm
D (21)	SO_2 in Regenerator Gas, μ amp Meter
D (22)	CO_2 in Regenerator Gas, μ amp Meter
D (23)	CO_2 in Boiler Gas, μ amp, Meter
D (24)	Spare
D (25)	Spare

Programme to File CAFB Pilot Plant Data
Basic Language - G-265 System

```
JHTPPF


100 REM JHTPPF FILES CAFB PILOT PLANT DATA
110 REM REVISED 17/12/71 TO USE MULTIPLE FILES
120 DATA 7
130 REM NUMBER OF DATA SETS TO BE ENTERED.  CHANGE EACH RUN.
140 FILES JHTFI1;JHTFI2;JHTFI3;JHTFI4;JHTFI5;JHTFI6
150 SET:1,1023
160 READ:1,H
170 RESTORE:H
180 READ:H,P
190 LET P1 = P+25
200 IF P1>1001 THEN 390
210 LET S = P + 1
220 LET H1 = H
230 SET:H,P+1
240 READ N
250 REM NUMBER OF DATA SETS TO BE PROCESSED THIS TIME
260 FOR J = 1 TO N
270 FOR I = 1 TO 25
280 READ D
281 IF I = 1 THEN 285
282 GO TO 300
285 PRINT D,P+1
290 REM D IS EACH OF THE DATA ITEMS TO BE READ IN
300 WRITE:H,D
310 LET P = P + 1
320 NEXT I
330 LET P1 = P+25
340 IF P1 > 1001 THEN 440
350 NEXT J
360 RESTORE: H
370 WRITE:H, P
380 GO TO 510
390 PRINT "FILE NO "; H; " IS FULL"
400 LET H = H+1
410 SET:1,1023
420 WRITE:1,H
430 GO TO 170
440 PRINT "FILE NO ";H; " IS FULL"
450 LET H = H+1
460 SET:1,1023
470 WRITE:1, H
480 LET P = 1
490 SET:H, 2
500 GO TO 350
510 PRINT "END OF DATA SET"
520 PRINT"THE NEXT DATA LOCATION IS FILE ";H;" LOCATION"; P+1
600 END
```

2. CAFB Velocities, Ratios, and Compositions

Programme JHT PPP uses data in files to compute various parameters of interest in CAFB operations. It can be run during the pilot plant operation to obtain a print out of recent operating conditions and it can be used to prepare summary tables at the end of a run. The current version of the programme prepares two tables. The first lists boiler and regenerator gas compositions computed from the analyser meter readings as well as sulphur removal efficiency in the gasifier and selectivity of CaS oxidation to CaO in the regenerator. These values are computed from feed rates and gas compositions. The second table lists gasifier superficial air velocity, the air/fuel ratio (percent of stoichiometric), the calcium/sulphur feed ratio, and regenerator SO_2 output as pounds of sulphur per hour and as percent of sulphur fed.

Programme to Compute CAFB Pilot Plant Gas Compositions, Efficiencies, Feed Ratios, and Regenerator Performance.

Basic Language - G265 System

```
JHTPPP


100 REM JHTPPP PROGRAM CALCULATES CAFB RESULTS FROM DATA IN FILES
110 REM REVISED 17/12/71 TO USE MULTIPLE FILES
120 DATA 86, 2.48
130 REM PCT CARBON AND SULFUR IN FUEL OIL
140 DATA 237
150 REM NUMBER OF DATA SETS TO BE ENTERED.  CHANGE EACH RUN.
160 FILES JHTFI1;JHTFI2;JHTFI3;JHTFI4;JHTFI5;JHTFI6
170 DIM D(25)
180 DIM A(10)
190 READ C1,C2
200 READ N
210 LET H1 = 1
220 PRINT USING 1560
230 PRINT
240 PRINT
250 PRINT USING 1570
260 PRINT USING 1580
270 PRINT USING 1590
280 PRINT USING 1600
290 PRINT
300 PRINT
310 LET K = 0
320 LET H = H1
330 SET:H, 2
340 FOR L = 1 TO N
350 FOR I = 1 TO 25
360 READ:H, D(I)
370 NEXT I
380 REM COMPUTE GAS COMPOSITIONS
390 REM  REGENERATOR CO2
400 LET Z2 = (D(22)/10) - 1
410 LET A2 = 1.1608245E-03 +1.1574973*Z2-.27330138*Z2↑2
420 LET A2 = A2+.075628236*Z2↑3-.00837005*Z2↑4+.00036410274*Z2↑5
430 REM REGENERATOR SO2
440 LET Z3 = (D(21)/10)-1
450 LET A3=.012783171+3.1254254*Z3-.63157951*Z3↑2
460LET A3 = A3+.14834157*Z3↑3-.015486024*Z3↑4+6.1794915E-04*Z3↑5
470 REM BOILER CO2
480LET Z4 = (D(23)/10)-1
490 LET A4=.01418178+2.321528*Z4-.51096758*Z4↑2
500LETA4=A4+.13242664*Z4↑3-.013988354*Z4↑4+6.0512861E-04*Z4↑5
510 LET F = D(13)*9.15
520 LET B1 = (1200/32)*C2*F
530 LET B2 = C1*F/100
540 LET B3 = 6*D(12)
550 LET B4 = B1*A4/(B2+B3)
560 REM B4 = PPM SO2 IN BOILER GAS IF NONE REMOVED
570 LET A(1) = 100 *(1-D(20)/B4)
580 LET A(2) = A4
590 LET A(3) = D(20)
```

Programme to Compute CAFB Pilot Plant Gas Compositions,
Efficiencies, Feed Ratios, and Regenerator Performance.
Basic Language G265 System

```
JHTPPP CONTINUED


600 LET Y1 = D(19)/100
610 LET Y2 = A2/100
620 LET Y3 = A3/100
630 LET D(11) = D(11)/60
640 LET A = D(11)*(406.9+D(9))/406.9
650 LET D(15) = D(15)/60
660 LET N1= D(15)*(406.9+D(10))/406.9
670 LET X1 = .21*A/(A+N1)
680 LET G = (1-X1)/(1-Y1-Y2-Y3)
690 LET C = ((X1/G)-Y1-Y2-(3/2)*Y3)/(2*Y3)
700 LET A(4) = 100*(1-(C/(C+1)))
710 LET A(5) = D(19)
720 LET A(6) = A2
730 LET A(7) = A3
740 LET A(8) = D(1)
750 LET A(9) = D(18)
760 LET A(10) = Y3*G*(A+N1)*(60*32/360)
770 REM SULPHUR OUTPUT OF REGENERATOR
780 SET:H, LOC(H)-2
790 WRITE:H, A(10)
800 LET W1 = F*C2/100
810 LET R1 = A(10)*100/W1
820 WRITE:H, R1
830 REM A(10)= S FROM REGEN; R1= PCT OF TOTAL S OUT REGEN
840 PRINT USING 1610,A(8),A(9),A(2),A(3),A(5),A(6),A(7),A(1),A(4)
850 LET K = K+1
860 IF K > 40 THEN 880
870 GO TO 1010
880 FOR P2 = 1 TO 20
890 PRINT
900 NEXT P2
910 LET K=0
920 PRINT USING 1560
930 PRINT
940 PRINT
950 PRINT USING 1570
960 PRINT USING 1580
970 PRINT USING 1590
980 PRINT USING 1600
990 PRINT
1000 PRINT
1010 IF LOC(H) > 1001 THEN 1030
1020 GO TO 1050
1030 LET H = H+1
1040 SET:H, 2
1050 NEXT L
1060 FOR P2 = 1 TO 20
1070 PRINT
1080 NEXT P2
1090 PRINT USING 1620
```

Programme to Compute CAFB Pilot Plant Gas Compositions, Efficiencies, Feed Ratios, and Regenerator Performance.

Basic Language - G265 System

```
JHTPPP CONTINUED


1100 PRINT
1110 PRINT
1120 PRINT USING 1630
1130 PRINT USING 1640
1140 PRINT
1150 PRINT
1160 LET F = D(13) * 9.15
1170 LET F1 = (60*100*29)/(360*13.92)
1180 LET F2=4.75886E-03
1190 LET F3=30.7429/C2
1200 LET K=0
1210 LET H = H1
1220 SET:H, 2
1230 FOR L = 1 TO N
1240 FOR I = 1 TO 25
1250 READ:H, D(I)
1260 NEXT I
1270 LET A5 = D(4) +(D(17)/21)*D(5)+ D(6)
1280 LET F = D(13) * 9.15
1290 LET A(1) = (F1*A5/F)
1300 LET A6 = D(4) + D(5) + D(6)
1310 LET A(2) = F2*A6*(D(2)+273)/(D(8)+  406.9)
1320 LET A(3) = F3*D(16)/F
1330 PRINT USING 1650,D(1),A(2),A(1),A(3),D(24),D(25)
1340 LET K = K+1
1350 IF K>40 THEN 1370
1360 GO TO 1470
1370 FOR P2 = 1 TO 20
1380 PRINT
1390 NEXT P2
1400 LET K=0
1410 PRINT USING 1620
1420 PRINT
1430 PRINT
1440 PRINT USING 1630
1450 PRINT USING 1640
1460 PRINT
1470 IF LOC(H) > 1001 THEN 1490
1480 GO TO 1510
1490 LET H = H+1
1500 SET:H, 2
1510 NEXT L
1520 FOR P2 = 1 TO 5
1530 PRINT
1540 NEXT P2
1550 PRINT "END OF DATA SET"
1560:                   CAFB GAS COMPOSITIONS AND EFFICIENCIES
1570:  TIME         BOILER GAS COMP     REGENERATOR GAS COMP
1580:               O2       CO2      SO2       O2       CO2      SO2  SULPHUR REGEN
1590:               VOL      VOL      PPM      VOL      VOL      VOL  REMOVAL CAO/CA
```

Programme to Compute CAFB Pilot Plant Gas Compositions,
Efficiencies, Feed Ratios, and Regenerator Performance.
Basic Language - G265 System

```
JHTPPP CONTINUED


1600: DAY.HOUR   PCT    PCT           PCT    PCT    PCT   PERCENT PERCEN
1610: ###.#### ###.#  ###.#  #####  ###.#  ###.#  ###.#   ####.#  ####.#
1620:          CAFB FEED RATIOS AND REGENERATOR SULPHUR
1630:DAY.HOUR     VELOCITY   AIR/FUEL  CA/S RATIO    REGENERATOR SULPHUR
1640:             FT/SEC    PCT STOIC  MOL/MOL       LB/HR   PCT OF FED
1650:###.####      ##.##      ###.#      ##.##       ###.##      ###.#
1660 END
```

3. Circulation Rates of Solids

Programme JHT PPW was prepared to estimate the circulation rate of solids from gasifier to regenerator on the basis of a heat balance using measured temperatures and gas compositions.

From gas compositions and flow rates the programme computes inlet and outlet gas enthalpies. The amount of CaS oxidised to $CaO + SO_2$ is determined from SO_2 production and used to estimate heat produced by this reaction. The measured CO_2 concentrations allows computation of heat release by carbon burning. These compositions, together with oxygen disappearance allow calculation of the amount of CaS converted to $CaSO_4$ and the heat released by that reaction. Heat losses through the regenerator wall are estimated from wall temperature. The programme then computes solids circulation rate by the equation,

$$\text{Inert Solids Rate} = \frac{\sum \text{Heat in} - \sum \text{Heat out}}{\text{Enthalpy of Solids out} - \text{Enthalpy of Solids in}}$$

$$\sum \text{of Heat in} = \sum \begin{array}{l} \text{Entering Gas Mass x Enthalpy} \\ \text{Entering Sulphur+Carbon Mass x Enthalpy} \\ \text{Heat of CaS Oxidation to } CaO + CaSO_4 \\ \text{Heat of Carbon oxidation to } CO_2 \end{array}$$

$$\sum \text{of Heat out} = \sum \begin{array}{l} \text{Heat transferred through walls} \\ \text{Exit gas Mass x Enthalpy} \end{array}$$

Table J-VI below lists the nomenclature of principal variables in the programme.

Since the programme assumes a steady state heat balance and makes no allowance for heat accumulation in the reactor walls or bed itself, results computed during periods of rapidly changing regenerator temperature are not valid. Where solids samples were available, circulation rates were calculated from sulphur balance measurements and compared with the heat balance estimates in Figure J - 30. Reasonable agreement was obtained.

TABLE J-VI

Nomenclature in Programme to Compute Solids Circulation Rates by Heat Balance on Regenerator

Variable (Starting with statement 750)	Meaning	
Z1	Off Gas	Mol/hr
Z2	Outlet CO_2	"
Z3	" SO_2	"
Z4	" O_2	"
Z5	" N_2	"
Z6	" $CaSO_4$	"
Z7	Inlet CaS	"
Z8	" Air	lb/hr
Z9	" N_2	"
T1	Gasifier Temp	Deg K
T2	Regenerator Temp	" "
T3	Gasifier Temp	Deg F
T4	Regenerator Temp	" "
T	Inlet Air Temp	" "
H2	Inlet Air Enthalpy	Btu/lb
H3	" N_2 "	"
H4	Outlet N_2 "	"
H5	" O_2 "	"
H6	" CO_2 "	"
H7	" SO_2 "	"
H8	Inlet CaO "	"
H9	Outlet CaO "	"
M1	Outlet $CaSO_4$ Enthalpy	Btu/lb
M2	Inlet Carbon "	"
M3	Heat Released by CaS ⟶ $CaSO_4$	Btu/hr
M4	Heat released by CaS ⟶ CaO	"
M5	Heat in with Air + N_2	"
M6	" " " CaS + Carbon	"
M7	Heat out with Off Gas	"
M8	Heat released by C ⟶ CO_2	"
M9	Heat Loss from Walls	"
Q1	Enthalpy Gain of Circulating CaO	Btu/lb
Q2	Heat inputs except from Circulating CaO	"
Q3	Heat outputs except from Circulating CaO	"
Q4	Heat Available to heat Circulating CaO	"
W1	CaO Circulation	lb/hr
W2	Solids Entering Regenerator	"
W3	Solids Exit Regenerator	"

Programme to Compute CAFB Pilot Plant Solids
Circulation Rates from Heat Balance
Basic Language G-265 System

```
JHTPPW


100 DATA 86,2.48
110 DATA 250
120 DATA -18.21464,.24642255,-2.5881903E-07,7.6981252E-09
130 DATA -1.5706847E-12,-15.982317,.25671377,-2.4932232E-05
140 DATA 3.0113785E-08,-6.942916E-12,-12.452016,.20982979
150 DATA 2.8205499E-05,-3.6796678E-09,-3.4144662E-13,-12.87373
160 DATA .19510153,7.2269052E-05
170 DATA -2.554598E-08,4.2340585E-12,-9.4090856,.14176497
180 DATA 3.4161995E-05,-7.7396955E-09,9.5608306E-13,-111.5,.322
190 DATA 7.79E-05,3470,-90,.2455,1.45E-04,2075,-193,.4,1.96E-04
200 DATA 17500,396000,197000,72,12,44,64,32,28,108000,2.75,1162,136
210 FILES JHTFI1;JHTFI2;JHTFI3;JHTFI4;JHTFI5;JHTFI6
220 DIM D(25)
230 DIM A(50)
240 READ C1,C2
250 READ N
260 FOR I = 1 TO 49
270 READ A(I)
280 NEXT I
290 PRINT USING 1350
300 PRINT
310 PRINT
320 PRINT USING 1360
330 PRINT USING 1370
340 PRINT USING 1380
350 PRINT USING 1390
360 PRINT
370 PRINT
380 LET H1 = 1
390 LET K = 0
400 LET H = H1
410 SET:H, 2
420 FOR L = 1 TO N
430 FOR I = 1 TO 25
440 READ:H, D(I)
450 NEXT I
460 REM COMPUTE GAS COMPOSITIONS
470 REM  REGENERATOR CO2
480 LET Z2 = (D(22)/10) - 1
490 LET A2 = 1.1608245E-03 +1.1574973*Z2-.27330138*Z2↑2
500 LET A2 = A2+.075628236*Z2↑3-.00837005*Z2↑4+.00036410274*Z2↑5
510 REM REGENERATOR SO2
520 LET Z3 = (D(21)/10)-1
530 LET A3=.012783171+3.1254254*Z3-.63157951*Z3↑2
540LET A3 = A3+.14834157*Z3↑3-.015486024*Z3↑4+6.1794915E-04*Z3↑5
550 REM BOILER CO2
560LET Z4 = (D(23)/10)-1
570 LET A4=.014181778+2.321528*Z4-.51096758*Z4↑2
580LETA4=A4+.13242664*Z4↑3-.013988354*Z4↑4+6.0512861E-04*Z4↑5
590 LET F = D(13)*9.15
```

Programme to Compute CAFB Pilot Plant Solids Circulation Rates from Heat Balance
Basic Language G-265 System

```
JHIPPW CONTINUED


600 LET B1 = (1200/32)*C2*F
610 LET B2 = C1*F/100
620 LET B3 = 6*D(12)
630 LET B4 = B1*A4/(B2+B3)
640 REM B4 = PPM SO2 IN BOILER GAS IF NONE REMOVED
650 LET Y1 = D(19)/100
660 LET Y2 = A2/100
670 LET Y3 = A3/100
680 LET D(11) = D(11)/60
690 LET A = D(11)*(406.9+D(9))/406.9
700 LET D(15) = D(15)/60
710 LET N1= D(15)*(406.9+D(10))/406.9
720 LET X1 = .21*A/(A+N1)
730 LET G = (1-X1)/(1-Y1-Y2-Y3)
740 LET C = ((X1/G)-Y1-Y2-(3/2)*Y3)/(2*Y3)
750 LET Z1 = G*(A+N1)*60/387
760 LET Z2 = Y2*Z1
770 LET Z3 = Y3*Z1
780 LET Z4 = Y1*Z1
790 LET Z5 = Z1-Z2-Z3-Z4
800 LET Z6 = Z3*C
810 LET Z7 = Z3+Z6
820 LET Z8 = 60 *A*29/387
830 LET Z9 = 60*N1*28/387
840 LET T1 = D(2)+273
850 LET T2 = D(3)+273
860 LET T3 = D(2)*1.8+32
870 LET T4 = D(3)*1.8+32
880 LET T = 80
890 LET H2=A(1)+A(2)*T+A(3)*T↑2+A(4)*T↑3+A(5)*T↑4
900 LET H3=A(6)+A(7)*T+A(8)*T↑2+A(9)*T↑3+A(10)*T↑4
910 LET H4=A(6)+A(7)*T4+A(8)*T4↑2+A(9)*T4↑3+A(10)*T4↑4
920 LET H5=A(11)+A(12)*T4+A(13)*T4↑2+A(14)*T4↑3+A(15)*T4↑4
930 LET H6 =A(16)+A(17)*T4+A(18)*T4↑2+A(19)*T4↑3+A(20)*T4↑4
940 LET H7=A(21)+A(22)*T4+A(23)*T4↑2+A(24)*T4↑3+A(25)*T4↑4
950 LET H8=A(26)+A(27)*T1+A(28)*T1↑2+A(29)/T1
960 LET H9=A(26)+A(27)*T2+A(28)*T2↑2+A(29)/T2
970 LET M1=A(30)+A(31)*T2+A(32)*T2↑2+A(33)/T2
980 LET M2=A(34)+A(35)*T2+A(36)*T2↑2+A(37)/T2
990 LET M3=Z6*A(38)
1000 LET M4=Z3*A(39)
1010 LET M5 = Z8*H2+Z9*H3
1020 LET M6=Z7*A(40)*H8+Z2*A(41)*M2
1030 LET M7=Z2*A(42)*H6+Z3*A(43)*H7+Z4*A(44)*H5+Z5*A(45)*H4
1040 LET M8=Z2*A(46)
1050 LET M9=A(47)*(T4-A(48))
1060 LET Q1 = H9-H8
1070 LET Q2 = M3+M4+M5+M6+M8
1080 LET Q3 = M7+M9
1090 LET Q4 = Q2-Q3
```

Programme to Compute CAFB Pilot Plant Solids Circulation Rates from Heat Balance
Basic Language G-265 System

JHTPPW CONTINUED

```
1100 LET W1 = Q4/Q1
1110 LET W2=W1+A(41)*Z2+A(40)*Z7
1120 LET W3 =W1+A(49)*Z6
1130 PRINT USING 1400,D(1), W1,W2,W3
1140 LET K=K+1
1150 IF K>40 THEN1170
1160 GO TO 1300
1170 FOR P2 = 1 TO 20
1180 PRINT
1190 NEXT P2
1200 LET K=0
1210 PRINT USING 1350
1220 PRINT
1230 PRINT
1240 PRINT USING 1360
1250 PRINT USING 1370
1260 PRINT USING 1380
1270 PRINT USING 1390
1280 PRINT
1290 PRINT
1300 IF LOC(H) > 1001 THEN 1320
1310 GO TO 1340
1320 LET H = H+1
1330 SET:H, 2
1340 NEXT L
1350:                CAFB SOLIDS CIRCULATION RATES
1360:     FLOW OF HOT SOLIDS IN POUNDS PER HOUR
1370:     TIME     CAO CIRCULATION      TOTAL SOLIDS FLOW
1380:   DAY.HOUR                        FROM         TO
1390:                                GASIFIER       GASIFIER
1400:   ###.####        #####         #####          #####
1410 END
```

4. Support Programmes

Several other simple programmes were written to assist in data editing, listing, and handling. In addition considerable use has been made of a G.265 system programme for data plotting on the teletype. This latter programme was used in preparing the hourly log plots of Run 3 data and results.

Support programmes include the following:

a. JHTTEM - Lists selected groups of original data from files.

b. JHT 99 - Initialises files for starting new data additions at first location.

c. JHT RD - Prints initial data items (Date.Hour) of each data set.

d. JHT COR - Permits correcting individual data items in files.

G. Data Tables, CAFB Pilot Plant, Run 3

TABLE J-VII

Composition of CAFB Solids Samples Run 3

Sample Time Day	Sample Time Hour	Sample Location	Sulphur	Composition, wt %		
				Sulphate Sulphur	Carbon	Vanadium
1	- 1730	Regenerator	3.40	0.95	0.0	0.52
1	- 1730	Gasifier Upper	7.61	0.09	0.13	0.62
2	- 1900	Regenerator	3.24	0.57	0.01	0.65
2	- 1900	Gasifier Upper	5.45	0.19	0.36	0.78
2	- 1900	Gasifier Lower	5.31	0.25	0.26	0.92
3	- 0230	Regenerator	3.17	0.64	0.0	0.93
3	- 0230	Gasifier Upper	6.15	0.15	0.31	0.72
3	- 0230	Gasifier Lower	5.93	0.20	0.27	0.82
3	- 0430	Gasifier Lower	5.06	0.14	0.37	1.15
3	- 0715	Gasifier Upper	5.08	0.13	0.45	1.15
3	- 0730	Regenerator	3.05	0.65	0.0	0.92
4	- 1430	Regenerator	3.16	0.88	0.04	0.95
5	- 0800	Gasifier Upper	3.91	0.19	0.06	1.1
5	- 0800	Gasifier Lower	3.86	0.26	0.05	1.5
5	- 0830	Regenerator	2.71	0.84	0.0	0.7
7	- 1130	Regenerator	2.37	1.18	0.02	1.05
7	- 1130	Gasifier Upper	4.15	0.23	0.15	1.5
7	- 1130	Gasifier Lower	4.10	0.27	0.12	1.5
8	- 0730	Gasifier Upper	3.65	0.19	0.22	1.4
8	- 0730	Gasifier Lower	3.86	0.20	0.23	1.45
9	- 0100	Gasifier Upper	4.81	0.32	0.01	1.4
9	- 0100	Gasifier Lower	5.47	0.35	0.01	0.97
9	- 2130	Regenerator	3.50	2.55	0.10	1.15
10	- 0530	Gasifier Upper	5.35	0.24	0.0	1.2
10	- 0530	Gasifier Lower	4.96	0.26	0.18	1.8
10	- 1230	Gasifier Upper	5.50	0.38	0.12	1.4
10	- 1230	Gasifier Lower	5.17	0.30	0.09	1.7

TABLE J-VIII

Composition of CAFB Gasifier Product Gas Samples, Run 3

Sample Time Day Hour	4-0500	7-1130	8-2230
Composition, vol %			
O_2	0	0	0
N_2	54.2	57.8	63.8
CO_2	10.3	9.4	11.4
CO	13.0	9.0	6.5
H_2	10.2	9.7	7.0
CH_4	5.5	6.0	4.8
C_2H_4	4.6	4.1	3.7
C_2H_6	0.4	0.2	0.3
C_3H_6	0.02	<0.01	<0.01
C_3H_8	0.09	0.2	0.1
C_4	0.08	0.3	0.09

TABLE J-IX

Sheet 1 of 20

Raw Data from Run 3

Day:Hour	Gasifier Temp. °C	Regenerator Temp °C	Gasifier: Air to Blower (SCFM)	Gasifier: Flue Gas Recycle (SCFM)	Gasifier: Air to Fuel Injectors + Start-up Burner (SCFM)
1.0230	875	895	170	70	18.0
1.0330	880	975	170	65	18.0
1.0430	900	980	220	55	18.0
1.0530	889	1011	210	65	19.3
1.0630	880	1085	200	60	19.3
1.0730	881	1108	200	60	18.0
1.0830	878	1088	200	60	18.0
1.0930	870	1088	200	63	18.0
1.1030	872	980	200	63	18.0
1.1130	890	1003	200	68	18.0
1.1230	890	1050	210	68	18.0
1.1330	900	1040	190	71	18.0
1.1430	905	1032	205	71	18.0
1.1530	885	1052	220	51	15.8
1.1630	882	1060	220	60	15.9
1.1730	888	1105	215	54	15.9
1.1830	888	1087	215	54	15.9
1.1930	898	1085	205	60	15.6
1.2030	880	1082	210	60	15.9
1.2130	875	1063	210	55	15.6
1.2230	860	1100	210	55	15.6
1.2330	882	1088	205	50	18.0
2.0030	870	1087	210	50	18.0
2.0130	908	1090	220	50	18.0
2.0230	868	1080	215	50	18.0
2.0330	872	1085	215	50	18.0
2.0430	875	1088	215	50	18.0
2.0530	867	1044	215	50	15.0
2.0630	862	1052	215	50	15.0
2.0730	869	1032	215	50	15.0
2.0830	870	1080	215	50	15.0
2.0930	875	1085	215	50	15.0
2.1030	878	1088	215	50	15.0
2.1130	868	1095	215	50	15.0
2.1230	872	1100	215	50	15.0
2.1330	878	1098	215	50	15.0
2.1430	875	1070	215	50	15.0
2.1530	882	1078	215	50	15.0
2.1630	882	1088	215	50	15.0
2.1730	886	1095	210	50	15.0
2.1830	882	1096	205	50	15.0

TABLE J-IX

Sheet 2 of 20

Raw Data from Run 3

Day:Hour	Gasifier Temp. °C	Regenerator Temp °C	Gasifier		
			Air to Blower (SCFM)	Flue Gas Recycle (SCFM)	Air to Fuel Injectors + Start-up Burner (SCFM)
2.1930	888	1098	205	50	15.0
2.2030	898	1100	205	50	15.0
2.2130	902	1098	200	50	15.0
2.2230	910	1097	210	50	15.0
2.2330	900	1098	200	50	15.0
3.0030	870	1098	195	50	15.0
3.0130	880	1098	200	50	15.0
3.0230	870	1090	210	50	15.0
3.0330	880	1095	200	50	15.0
3.0430	885	1100	200	50	15.0
3.0530	865	1095	200	50	15.0
3.0630	870	1098	200	50	15.0
3.0730	870	1098	200	50	15.0
3.0830	870	1098	190	50	15.0
3.0930	865	1085	185	50	15.0
3.1030	865	1095	190	50	15.0
3.1130	870	1095	195	50	15.0
3.1230	877	1095	190	35	13.8
3.1330	885	1100	190	35	13.8
3.1430	880	1095	190	40	13.8
3.1630	898	1098	190	45	20.1
3.1730	909	1082	180	45	31.5
3.2305	880	1090	190	50	21.0
4.0005	880	1092	210	55	21.0
4.0130	880	1090	200	65	33.0
4.0230	890	1030	210	60	33.0
4.0330	886	1020	210	60	33.0
4.0430	880	1025	210	55	33.0
4.0530	910	1045	210	55	33.0
4.0630	899	1050	200	60	30.0
4.0730	885	1095	195	60	30.0
4.0830	900	1065	200	60	30.0
4.0930	910	1084	210	55	30.0
4.1030	890	1092	210	55	30.0
4.1130	888	1039	205	55	25.6
4.1230	915	1085	210	55	19.3
4.1330	910	1092	205	55	18.9
4.1430	910	1098	205	55	19.1
4.1530	912	1095	200	55	20.4
4.1630	910	1081	200	50	20.8
4.1730	915	1095	200	52	18.4

TABLE J-IX

Sheet 3 of 20

Raw Data from Run 3

Day:Hour	Gasifier Temp. °C	Regenerator Temp °C	Gasifier Air to Blower (SCFM)	 Flue Gas Recycle (SCFM)	 Air to Fuel Injectors + Start-up Burner (SCFM)
4.1830	898	1098	195	55	18.6
4.1930	902	1094	200	50	17.9
5.0130	900	980	210	60	21.0
5.0230	895	1010	205	60	21.0
5.0330	900	1070	200	60	18.0
5.0430	890	1090	200	60	18.0
5.0530	885	1085	200	60	18.0
5.0630	885	1090	200	60	18.0
5.0730	910	1095	200	60	18.0
5.0930	890	1092	190	60	18.0
5.1030	920	1095	210	50	18.0
5.1130	928	1080	200	50	18.0
5.1230	920	1098	195	55	18.0
5.1330	920	1098	190	55	15.0
5.1430	920	1094	185	55	15.0
5.1530	905	1089	180	60	15.0
5.1630	880	1130	150	75	15.0
5.1730	860	1060	150	70	-
5.2200	925	1015	190	70	15.9
5.2230	927	1084	175	75	18.0
5.2330	855	1010	195	54	16.5
6.0030	920	1058	195	40	15.0
6.0130	885	1095	180	40	18.0
6.0230	895	1095	160	50	18.0
6.0330	900	1092	165	55	18.0
6.0430	875	1095	165	55	18.0
6.0530	883	1098	165	55	18.0
6.0630	882	1098	160	60	18.0
6.1130	961	780	200	65	14.5
6.1230	-	-	210	55	14.5
6.1330	-	-	210	50	11.2
6.1430	-	-	165	50	11.5
6.1500	913	1104	185	50	11.4
6.1830	-	-	240	60	19.0
6.1930	-	-	220	70	18.0
6.2030	883	1088	190	85	18.0
6.2130	860	1088	190	92	18.0
6.2230	882	1090	205	80	18.0
6.2330	900	1098	205	80	18.0
7.0030	883	1095	210	77	18.0
7.0130	883	1100	210	77	18.0

TABLE J-IX

Sheet 4 of 20

Raw Data from Run 3

Day:Hour	Gasifier Temp. °C	Regenerator Temp °C	Gasifier Air to Blower (SCFM)	Gasifier Flue Gas Recycle (SCFM)	Gasifier Air to Fuel Injectors + Start-up Burner (SCFM)
7.0230	890	1100	210	77	18.0
7.0330	888	1100	210	79	18.0
7.0430	883	1098	210	79	18.0
7.0530	901	1098	210	77	18.0
7.0630	880	1098	210	77	18.0
7.0730	885	1100	205	77	18.0
7.0830	870	1100	205	77	18.0
7.0930	870	1098	200	75	17.0
7.1030	878	1099	200	77	17.2
7.1130	880	1093	200	78	18.0
7.1230	897	1090	200	77	18.0
7.1330	888	1092	200	80	18.0
7.1430	888	1090	195	125	18.0
7.1530	887	1089	195	85	18.0
7.1630	890	1090	195	85	17.8
7.1730	884	1091	190	87	17.8
7.1830	-	-	190	90	17.8
7.1930	885	1090	190	90	17.8
7.2030	875	1092	190	95	16.2
7.2130	872	1090	180	95	16.3
7.2230	867	1097	180	95	16.5
7.2330	868	1100	180	95	16.2
8.0030	863	1108	180	95	20.8
8.0130	858	1100	180	95	20.8
8.0230	874	1100	180	90	18.9
8.0330	862	1100	190	90	18.4
8.0430	864	1103	190	90	19.2
8.0530	855	1090	190	90	17.8
8.0630	863	1070	190	90	17.9
8.0730	863	1070	190	85	18.0
8.0830	874	1081	190	87	19.0
8.0930	882	1068	190	87	19.0
8.1130	880	1065	190	80	19.0
8.1230	872	1068	190	82	19.0
8.1330	892	970	190	82	19.0
8.1430	863	1062	190	82	19.0
8.2030	928	870	180	120	18.0
8.2130	882	1055	200	100	18.0
8.2230	882	1058	200	110	18.0
8.2330	870	1068	180	100	18.0
9.0030	875	1070	180	110	18.0

TABLE J-IX

Sheet 5 of 20

Raw Data from Run 3

Day:Hour	Gasifier Temp. °C	Regenerator Temp °C	Gasifier		
			Air to Blower (SCFM)	Flue Gas Recycle (SCFM)	Air to Fuel Injectors + Start-up Burner (SCFM)
9.0130	878	1070	200	90	18.0
9.0230	882	1070	195	90	18.0
9.0330	878	1068	200	100	18.0
9.0430	882	1050	195	100	18.0
9.0530	880	1065	200	100	18.0
9.0630	885	1065	185	100	18.0
9.0730	868	1055	185	115	18.0
9.0830	868	1058	190	120	18.0
9.0930	878	1068	190	110	17.8
9.1030	878	1065	190	110	17.0
9.1130	868	1068	185	107	17.9
9.1530	880	720	190	80	17.6
9.1630	-	-	200	85	18.0
9.1730	880	1062	190	95	18.0
9.1830	872	1068	190	95	18.0
9.1930	888	1060	195	88	18.0
9.2030	860	1070	190	90	18.0
9.2130	858	1087	200	94	18.0
9.2230	881	1062	200	92	18.0
9.2330	869	1061	180	95	18.0
10.0030	869	1063	190	97	18.0
10.0130	866	1068	190	97	18.0
10.0230	871	1078	190	97	18.0
10.0330	877	1085	190	95	18.0
10.0430	881	1069	-	-	-
10.0530	876	1069	190	95	18.0
10.0630	889	1068	190	95	18.0
10.0730	901	1066	200	90	18.0
10.0830	885	1072	200	85	18.0
10.0930	875	1068	180	90	18.0
10.1030	875	1068	190	95	18.0
10.1130	882	1070	190	89	18.0
10.1230	864	1069	190	90	18.0
10.1330	868	1068	185	90	18.0
10.1430	868	1070	185	90	18.0
10.1530	872	1070	185	90	18.0
10.1630	850	1068	140	105	18.0

TABLE J-IX

Sheet 6 of 20

Raw Data from Run 3

Day:Hour	Gasifier Pressure (inch.W.G)	Regenerator Gas Meter Pressure (inches W.G)	N_2 Meter Pressure (inches W.G)	Regenerator Air (SCFH)	Propane to Pilot (SCFM)	Oil Rate (gall/hr)
1.0230	9.00	36.0	83.0	1020	2.0	33.0
1.0330	9.00	36.0	83.0	1100	2.0	33.0
1.0430	11.00	36.0	83.0	800	2.0	37.0
1.0530	11.00	36.0	83.0	700	2.0	41.5
1.0630	11.00	36.0	83.0	990	2.0	41.5
1.0730	11.25	36.0	83.0	910	2.0	41.5
1.0830	11.25	36.0	83.0	800	2.0	41.5
1.0930	11.20	36.0	83.0	870	2.0	41.5
1.1030	11.00	36.0	83.0	790	2.0	41.5
1.1130	10.80	36.0	83.0	880	2.0	41.5
1.1230	11.50	36.0	83.0	880	2.0	41.5
1.1330	12.00	36.0	83.0	800	2.0	41.5
1.1430	12.00	36.0	83.0	840	2.0	41.5
1.1530	11.50	36.0	83.0	951	2.0	41.5
1.1630	11.50	36.0	83.0	951	2.0	41.5
1.1730	12.00	36.0	83.0	1065	2.0	41.5
1.1830	12.00	36.0	83.0	1050	2.0	41.5
1.1930	11.50	36.0	83.0	1195	2.0	41.5
1.2030	12.00	36.0	83.0	1200	2.0	41.5
1.2130	11.75	36.0	83.0	1095	2.0	41.5
1.2230	12.50	36.0	83.0	1095	2.0	41.5
1.2330	12.50	36.0	83.0	1100	2.0	41.5
2.0030	12.80	38.0	83.0	1103	2.0	41.5
2.0130	13.00	38.0	83.0	1082	2.0	41.5
2.0230	13.00	38.0	83.0	1107	2.0	41.5
2.0330	13.00	38.0	83.0	1107	2.0	41.5
2.0430	12.80	38.0	83.0	1089	2.0	41.5
2.0530	12.80	38.0	83.0	1102	2.0	41.5
2.0630	12.80	38.0	83.0	1107	2.0	41.5
2.0730	13.00	38.0	83.0	1093	2.0	41.5
2.0830	13.00	38.0	83.0	1010	2.0	41.5
2.0930	12.50	38.0	83.0	915	2.0	41.5
2.1030	12.70	38.0	83.0	1210	2.0	41.5
2.1130	12.70	38.0	83.0	1210	2.0	41.5
2.1230	12.70	38.0	83.0	940	2.0	41.5
2.1330	12.70	39.0	83.0	1170	2.0	41.5
2.1430	12.70	39.0	83.0	1030	2.0	41.5
2.1530	12.70	39.0	83.0	830	2.0	41.5
2.1630	12.70	39.0	83.0	1130	2.0	41.5
2.1730	13.00	39.0	83.0	910	2.0	41.5
2.1830	13.10	39.0	83.0	1070	2.0	41.5

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Raw Data from Run 3

Day:Hour	Gasifier Pressure (inch.W.G)	Regenerator Gas Meter Pressure (inches W.G)	N_2 Meter Pressure (inches W.G)	Regenerator Air (SCFH)	Propane to Pilot (SCFM)	Oil Rate (gall/hr)
2.1930	13.10	39.0	83.0	850	2.0	41.5
2.2030	13.30	39.0	83.0	840	2.0	41.5
2.2130	13.80	39.0	83.0	895	2.0	41.5
2.2230	13.50	39.0	83.0	740	2.0	41.5
2.2330	13.50	39.0	83.0	910	2.0	41.5
3.0030	13.50	39.0	83.0	755	2.0	41.5
3.0130	13.50	40.0	83.0	870	2.0	41.5
3.0230	14.00	40.0	83.0	960	2.0	41.5
3.0330	14.00	40.0	83.0	1000	2.0	41.5
3.0430	14.00	40.0	83.0	1010	2.0	41.5
3.0530	14.00	40.0	83.0	1410	2.0	41.5
3.0630	14.00	40.0	83.0	990	2.0	41.5
3.0730	14.00	40.0	83.0	880	2.0	41.5
3.0830	14.50	40.0	83.0	1050	2.0	41.5
3.0930	14.50	40.0	83.0	1050	2.0	41.5
3.1030	14.50	40.0	83.0	1090	2.0	41.5
3.1130	14.70	41.0	83.0	960	2.0	41.5
3.1230	14.70	41.0	83.0	990	2.0	41.5
3.1330	15.00	41.0	83.0	1060	2.0	41.5
3.1430	15.00	41.0	83.0	980	2.0	41.5
3.1630	15.50	41.0	83.0	925	2.0	41.5
3.1730	16.50	41.0	83.0	750	2.0	41.5
3.2305	-	41.0	83.0	1020	2.0	41.3
4.0005	14.00	41.0	83.0	1130	2.0	41.3
4.0130	14.00	41.0	83.0	932	2.0	41.3
4.0230	14.00	41.0	83.0	1010	2.0	41.3
4.0330	14.00	41.0	83.0	950	2.0	41.3
4.0430	13.80	41.0	83.0	850	2.0	41.3
4.0530	13.50	41.0	83.0	1210	2.0	41.3
4.0630	13.50	42.0	83.0	1210	2.0	41.3
4.0730	14.50	42.0	83.0	1090	2.0	41.3
4.0830	14.00	42.0	83.0	980	2.0	41.3
4.0930	14.00	42.0	83.0	1270	2.0	41.3
4.1030	14.50	42.0	83.0	810	2.0	41.3
4.1130	14.00	42.0	83.0	920	2.0	41.3
4.1230	14.50	42.0	83.0	1020	2.0	41.3
4.1330	14.50	42.0	83.0	940	2.0	41.3
4.1430	14.50	42.0	83.0	930	2.0	41.3
4.1530	14.50	42.0	83.0	980	2.0	41.3
4.1630	15.00	42.0	83.0	920	2.0	41.3
4.1730	15.30	42.0	83.0	1070	2.0	41.3

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Raw Data from Run 3

Day:Hour	Gasifier Pressure (inch.W.G)	Regenerator Gas Meter Pressure (inches W.G)	N_2 Meter Pressure (inches W.G)	Regenerator Air (SCFH)	Propane to Pilot (SCFM)	Oil Rate (gall/hr)
4.1830	15.50	43.0	83.0	1042	2.0	41.3
4.1930	16.00	43.0	83.0	870	2.0	41.3
5.0130	13.00	43.0	83.0	-	2.0	39.4
5.0230	12.50	43.0	83.0	-	2.0	41.0
5.0330	13.00	43.0	83.0	920	2.0	40.0
5.0430	13.20	43.0	83.0	1000	2.0	39.0
5.0530	13.50	43.0	83.0	1020	2.0	40.0
5.0630	13.50	43.0	83.0	1020	2.0	40.7
5.0730	14.00	43.0	83.0	1010	2.0	39.0
5.0930	14.00	43.0	83.0	870	2.0	39.4
5.1030	14.50	43.0	83.0	981	2.0	39.4
5.1130	14.40	43.0	83.0	905	2.0	39.4
5.1230	15.00	44.0	83.0	1014	2.0	39.4
5.1330	15.30	44.0	83.0	990	2.0	39.4
5.1430	16.40	44.0	83.0	1001	2.0	39.4
5.1530	17.50	44.0	83.0	890	2.0	39.4
5.1630	18.50	44.0	83.0	895	2.0	39.4
5.1730	20.50	44.0	83.0	904	2.0	39.4
5.2200	14.80	44.0	83.0	-	2.0	39.2
5.2230	15.00	44.0	83.0	1010	2.0	39.2
5.2330	16.00	44.0	83.0	1021	2.0	39.2
6.0030	17.00	44.0	83.0	1004	2.0	39.2
6.0130	16.50	44.0	83.0	965	2.0	39.2
6.0230	17.00	44.0	83.0	912	2.0	39.2
6.0330	17.50	45.0	83.0	1101	2.0	39.2
6.0430	18.00	45.0	83.0	882	2.0	39.2
6.0530	18.00	45.0	83.0	1042	2.0	39.2
6.0630	19.10	45.0	83.0	1032	2.0	39.2
6.1130	14.00	45.0	83.0	810	2.0	42.0
6.1230	-	45.0	83.0	995	2.0	42.2
6.1330	-	45.0	83.0	1098	2.0	42.2
6.1430	-	45.0	83.0	1100	2.0	42.2
6.1500	18.70	45.0	83.0	610	2.0	42.0
6.1830	-	46.0	83.0	-	2.0	42.2
6.1930	-	46.0	83.0	660	2.0	42.2
6.2030	18.00	46.0	83.0	978	2.0	42.0
6.2130	16.00	46.0	83.0	928	2.0	42.0
6.2230	16.80	46.0	83.0	910	2.0	42.2
6.2330	17.00	46.0	83.0	921	2.0	42.2
7.0030	17.50	46.0	83.0	937	2.0	42.2
7.0130	17.50	46.0	83.0	848	2.0	42.2

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Raw Data from Run 3

Day:Hour	Gasifier Pressure (inch.W.G)	Regenerator Gas Meter Pressure (inches W.G)	N_2 Meter Pressure (inches W.G)	Regenerator Air (SCFH)	Propane to Pilot (SCFM)	Oil Rate (gall/hr)
7.0230	17.50	46.0	83.0	874	2.0	42.2
7.0330	17.50	46.0	83.0	850	2.0	42.2
7.0430	17.60	46.0	83.0	830	2.0	42.2
7.0530	18.10	46.0	83.0	824	2.0	42.2
7.0630	18.10	46.0	83.0	849	2.0	42.2
7.0730	18.00	46.0	83.0	859	2.0	42.2
7.0830	18.30	47.0	83.0	864	2.0	42.2
7.0930	18.30	47.0	83.0	847	2.0	42.2
7.1030	18.50	47.0	83.0	811	2.0	42.2
7.1130	18.80	47.0	83.0	782	2.0	42.2
7.1230	19.50	47.0	83.0	781	2.0	42.2
7.1330	19.70	47.0	83.0	857	2.0	42.2
7.1430	19.70	47.0	83.0	948	2.0	42.2
7.1530	19.80	47.0	83.0	942	2.0	42.2
7.1630	19.80	47.0	83.0	958	2.0	42.2
7.1730	19.80	47.0	83.0	970	2.0	42.2
7.1830	-	47.0	83.0	954	2.0	42.2
7.1930	20.50	47.0	83.0	954	2.0	42.2
7.2030	21.00	48.0	83.0	944	2.0	42.2
7.2130	21.50	48.0	83.0	936	2.0	42.2
7.2230	21.50	48.0	83.0	914	2.0	42.2
7.2330	21.00	48.0	83.0	901	2.0	42.2
8.0030	22.00	48.0	83.0	882	2.0	42.2
8.0130	21.50	48.0	83.0	881	2.0	42.2
8.0230	22.00	48.0	83.0	901	2.0	42.2
8.0330	21.60	48.0	83.0	900	2.0	42.2
8.0430	22.00	48.0	83.0	953	2.0	42.2
8.0530	22.00	48.0	83.0	899	2.0	42.2
8.0630	22.00	48.0	83.0	933	2.0	42.2
8.0730	22.30	48.0	83.0	794	2.0	42.2
8.0830	22.30	49.0	83.0	782	2.0	33.0
8.0930	22.50	49.0	83.0	811	2.0	33.0
8.1130	21.50	49.0	83.0	823	2.0	33.0
8.1230	21.50	49.0	83.0	826	2.0	33.0
8.1330	21.80	49.0	83.0	824	2.0	33.0
8.1430	21.20	49.0	83.0	893	2.0	33.0
8.2030	16.00	49.0	83.0	-	2.0	38.9
8.2130	15.20	49.0	83.0	860	2.0	38.9
8.2230	15.50	49.0	83.0	820	2.0	38.9
8.2330	16.00	49.0	83.0	816	2.0	38.9
9.0030	15.50	49.0	83.0	814	2.0	38.9

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Raw Data from Run 3

Day:Hour	Gasifier Pressure (inch.W.G)	Regenerator Gas Meter Pressure (inches W.G)	N_2 Meter Pressure (inches W.G)	Regenerator Air (SCFH)	Propane to Pilot (SCFM)	Oil Rate (gall/hr)
9.0130	16.20	49.0	83.0	830	2.0	38.9
9.0230	16.20	50.0	83.0	812	2.0	38.9
9.0330	16.20	50.0	83.0	762	2.0	38.9
9.0430	18.00	50.0	83.0	1008	2.0	38.9
9.0530	18.00	50.0	83.0	1052	2.0	38.9
9.0630	18.00	50.0	83.0	971	2.0	38.9
9.0730	18.00	50.0	83.0	1000	2.0	38.9
9.0830	18.00	50.0	83.0	995	2.0	38.9
9.0930	18.50	50.0	83.0	992	2.0	38.9
9.1030	19.0	50.0	83.0	980	2.0	-
9.1130	19.00	50.0	83.0	972	2.0	38.9
9.1530	15.50	50.0	83.0	-	2.0	38.6
9.1630	18.00	50.0	83.0	952	2.0	38.6
9.1730	18.00	51.0	83.0	940	2.0	38.6
9.1830	18.30	51.0	83.0	984	2.0	38.6
9.1930	18.70	51.0	83.0	1022	2.0	38.6
9.2030	15.70	51.0	83.0	957	2.0	38.6
9.2130	19.10	51.0	83.0	867	2.0	38.6
9.2230	19.60	51.0	83.0	860	2.0	38.6
9.2330	19.90	51.0	83.0	830	2.0	38.6
10.0030	19.30	51.0	83.0	800	2.0	38.6
10.0130	19.50	51.0	83.0	810	2.0	38.6
10.0230	19.70	51.0	83.0	810	2.0	38.6
10.0330	19.50	51.0	83.0	860	2.0	38.6
10.0430	19.50	51.0	83.0	850	2.0	38.6
10.0530	19.70	52.0	83.0	850	2.0	38.6
10.0630	20.20	52.0	83.0	840	2.0	38.6
10.0730	19.80	52.0	83.0	850	2.0	38.6
10.0830	20.00	53.0	83.0	840	2.0	38.6
10.0930	20.20	52.0	83.0	857	2.0	35.0
10.1030	20.20	52.0	83.0	851	2.0	35.0
10.1130	20.30	52.0	83.0	855	2.0	35.0
10.1230	20.50	52.0	83.0	847	2.0	35.0
10.1330	20.50	52.0	83.0	846	2.0	35.0
10.1430	20.70	52.0	83.0	843	2.0	35.0
10.1530	21.00	52.0	83.0	831	2.0	35.0
10.1630	21.00	52.0	83.0	835	2.0	35.0

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Raw Data from Run 3

Day:Hour	Oil Temp (°F)	N_2 to Regenerator (SCFH)	Stone Feed Rate lb/hr	O_2 in Plenum Gas (% by vol.)	O_2 in Boiler Gas (% by vol).
1.0230	205	51	33.0	-	6.80
1.0330	200	51	33.0	-	6.90
1.0430	190	51	33.0	-	3.90
1.0530	175	51	33.0	-	3.40
1.0630	195	51	33.0	-	2.80
1.0730	185	51	33.0	-	3.90
1.0830	200	51	34.0	-	[illegible]
1.0930	185	51	34.0	-	[illegible]
1.1030	185	51	34.0	-	3.30
1.1130	195	51	34.0	-	3.30
1.1230	200	51	35.0	-	3.20
1.1330	185	51	34.0	-	[illegible]
1.1430	190	51	34.0	-	[illegible]
1.1530	190	47	47.0	-	[illegible]
1.1630	175	47	47.0	-	[illegible]
1.1730	210	47	47.0	-	[illegible]
1.1830	175	47	47.0	-	[illegible]
1.1930	175	47	47.0	-	3.10
1.2030	195	47	59.0	-	[illegible]
1.2130	175	47	59.0	-	[illegible]
1.2230	185	47	59.0	-	[illegible]
1.2330	200	47	59.0	-	3.30
2.0030	200	56	79.0	-	3.10
2.0130	200	56	79.0	-	3.20
2.0230	200	56	79.0	-	3.20
2.0330	200	56	104.0	-	3.30
2.0430	175	56	124.0	-	3.30
2.0530	190	56	66.0	-	3.40
2.0630	200	56	66.0	-	3.30
2.0730	200	56	66.0	-	3.30
2.0830	200	56	76.0	-	3.30
2.0930	190	56	76.0	-	3.30
2.1030	190	56	76.0	-	3.10
2.1130	205	56	59.0	-	3.50
2.1230	205	37	59.0	-	3.10
2.1330	200	37	59.0	-	3.50
2.1430	175	37	59.0	-	3.50
2.1530	200	37	45.0	-	3.20
2.1630	200	37	45.0	-	3.50
2.1730	205	36	45.0	-	3.20
2.1830	205	36	45.0	-	3.60

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Raw Data from Run 3

Day:Hour	Oil Temp (°F)	N_2 to Regenerator (SCFH)	Stone Feed Rate lb/hr	O_2 in Plenum Gas (% by vol.)	O_2 in Boiler Gas (% by vol).
2.1930	200	36	45.0	-	3.20
2.2030	200	36	45.0	-	3.40
2.2130	200	36	33.0	-	3.20
2.2230	190	36	33.0	-	3.50
2.2330	200	36	33.0	-	3.50
3.0030	200	36	33.0	-	3.10
3.0130	200	36	33.0	-	3.20
3.0230	210	39	33.0	-	3.50
3.0330	200	39	33.0	-	3.00
3.0430	200	39	33.0	-	3.50
3.0530	205	39	47.0	-	3.80
3.0630	210	39	47.0	-	3.80
3.0730	200	39	47.0	-	4.00
3.0830	200	39	47.0	-	3.50
3.0930	200	39	47.0	-	3.50
3.1030	200	39	53.0	-	4.00
3.1130	200	39	53.0	-	4.20
3.1230	200	39	53.0	-	3.50
3.1330	200	39	53.0	-	3.50
3.1430	200	39	31.0	-	5.80
3.1630	200	39	31.0	-	3.40
3.1730	200	39	31.0	-	3.50
3.2305	200	28	31.0	-	-
4.0005	200	28	31.0	-	3.50
4.0130	200	28	109.0	-	2.50
4.0230	210	28	109.0	-	2.90
4.0330	200	28	24.0	-	2.50
4.0430	200	28	24.0	-	2.50
4.0530	200	28	24.0	-	3.00
4.0630	200	28	24.0	-	2.50
4.0730	200	28	24.0	-	2.20
4.0830	200	28	24.0	-	3.10
4.0930	175	28	24.0	-	3.60
4.1030	200	28	24.0	-	3.60
4.1130	200	28	46.0	-	3.10
4.1230	200	28	46.0	-	3.50
4.1330	200	28	46.0	-	3.50
4.1430	190	28	46.0	-	3.00
4.1530	200	28	46.0	-	3.00
4.1630	200	28	41.0	-	3.00
4.1730	200	28	41.0	-	3.00

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Raw Data from Run 3

Day:Hour	Oil Temp (°F)	N_2 to Regenerator (SCFH)	Stone Feed Rate lb/hr	O_2 in Plenum Gas (% by vol.)	O_2 in Boiler Gas (% by vol).
4.1830	200	28	41.0	-	3.00
4.1930	200	28	41.0	-	3.00
5.0130	200	28	51.0	-	2.10
5.0230	200	28	51.0	-	3.00
5.0330	200	28	51.0	-	2.50
5.0430	200	28	64.0	-	3.00
5.0530	200	28	64.0	-	3.00
5.0630	200	28	64.0	-	3.00
5.0730	200	28	64.0	-	2.80
5.0930	180	28	37.0	-	2.10
5.1030	200	28	37.0	-	3.10
5.1130	200	28	37.0	-	3.10
5.1230	200	28	37.0	-	3.50
5.1330	185	28	37.0	-	-
5.1430	200	28	37.0	-	2.70
5.1530	180	28	37.0	-	2.90
5.1630	180	28	58.0	-	3.70
5.1730	180	28	58.0	-	3.10
5.2200	205	30	58.0	-	4.20
5.2230	175	30	58.0	-	2.80
5.2330	200	30	58.0	-	4.70
6.0030	190	30	16.0	-	5.70
6.0130	172	30	16.0	-	2.10
6.0230	200	30	16.0	-	3.40
6.0330	173	30	16.0	-	1.90
6.0430	175	30	16.0	-	3.00
6.0530	210	30	16.0	-	2.70
6.0630	175	30	16.0	-	1.60
6.1130	200	25	16.0	-	-
6.1230	180	25	16.0	-	-
6.1330	200	25	16.0	-	-
6.1430	200	25	16.0	-	-
6.1500	180	25	16.0	-	2.40
6.1830	200	35	56.0	-	-
6.1930	180	35	56.0	-	-
6.2030	200	35	56.0	-	2.40
6.2130	200	35	56.0	-	2.00
6.2230	200	35	56.0	-	2.00
6.2330	195	35	42.0	-	2.00
7.0030	200	35	42.0	-	2.00
7.0130	200	35	42.0	-	2.10

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Raw Data from Run 3

Day:Hour	Oil Tem. (°F)	N_2 to Regenerator (SCFH)	Stone Feed Rate lb/hr	O_2 in Plenum Gas (% by vol.)	O_2 in Boiler Gas (% by vol).
7.0230	213	35	42.0	-	2.00
7.0330	210	35	42.0	-	2.20
7.0430	215	35	42.0	-	2.00
7.0530	180	35	59.0	-	1.80
7.0630	200	35	59.0	-	2.00
7.0730	203	35	59.0	-	2.00
7.0830	193	35	59.0	-	2.20
7.0930	190	35	36.0	-	2.40
7.1030	200	35	36.0	14.5	2.30
7.1130	200	35	36.0	14.3	2.00
7.1230	200	35	36.0	14.5	2.00
7.1330	200	125	36.0	14.0	2.30
7.1430	200	125	36.0	13.8	2.10
7.1530	185	125	28.0	14.0	2.80
7.1630	210	125	28.0	13.7	-
7.1730	180	58	28.0	13.4	2.40
7.1830	195	58	28.0	13.4	-
7.1930	200	58	28.0	13.4	2.20
7.2030	200	58	28.0	13.2	1.80
7.2130	215	58	28.0	13.3	2.00
7.2230	215	58	28.0	13.5	2.00
7.2330	215	58	47.0	13.5	2.00
8.0030	215	58	47.0	13.0	2.20
8.0130	175	58	47.0	13.0	2.00
8.0230	195	58	47.0	13.8	2.70
8.0330	212	58	47.0	13.6	2.00
8.0430	200	58	64.0	13.4	2.00
8.0530	190	58	64.0	13.5	2.00
8.0630	195	58	64.0	13.7	2.00
8.0730	212	58	28.0	13.7	2.10
8.0830	210	58	28.0	13.8	1.90
8.0930	210	58	28.0	13.8	2.10
8.1130	214	58	28.0	13.8	3.00
8.1230	194	58	28.0	15.4	2.80
8.1330	198	58	28.0	16.0	6.00
8.1430	214	58	40.0	14.0	3.60
8.2030	204	60	40.0	13.8	5.20
8.2130	200	60	40.0	13.5	3.00
8.2230	200	60	40.0	14.0	3.10
8.2330	215	60	40.0	13.5	2.50
9.0030	200	60	52.0	13.8	2.50

TABLE J-IX

Raw Data from Run 3

Day:Hour	Oil Temp (°F)	N_2 to Regenerator (SCFH)	Stone Feed Rate lb/hr	O_2 in Plenum Gas (% by vol.)	O_2 in Boiler Gas (% by vol).
9.0130	200	60	52.0	13.8	2.70
9.0230	200	-	52.0	13.8	2.20
9.0330	200	50	52.0	13.5	2.30
9.0430	200	50	52.0	13.8	2.20
9.0530	200	50	30.0	13.8	2.00
9.0630	200	50	30.0	13.2	2.00
9.0730	195	50	30.0	13.2	2.10
9.0830	200	50	30.0	13.2	2.50
9.0930	200	50	30.0	13.8	2.20
9.1030	200	51	30.0	13.5	-
9.1130	190	50	30.0	13.6	2.50
9.1530	180	85	30.0	13.4	-
9.1630	200	85	26.0	13.4	-
9.1730	200	85	26.0	13.2	2.60
9.1830	200	85	26.0	13.3	2.10
9.1930	200	85	26.0	13.8	2.10
9.2030	200	85	26.0	13.8	1.80
9.2130	185	85	26.0	13.8	1.90
9.2230	210	85	26.0	13.0	1.50
9.2330	200	85	26.0	13.4	2.10
10.0030	200	85	26.0	13.2	2.30
10.0130	205	85	28.0	13.0	1.90
10.0230	185	85	28.0	13.0	2.10
10.0330	210	85	28.0	13.2	1.90
10.0430	-	85	28.0	-	1.80
10.0530	200	85	28.0	13.5	1.60
10.0630	200	100	28.0	13.4	1.70
10.0730	200	100	28.0	13.0	2.20
10.0830	200	100	15.0	13.5	2.00
10.0930	200	100	15.0	13.1	1.60
10.1030	200	71	15.0	13.0	1.60
10.1130	194	71	15.0	13.1	1.20
10.1230	200	71	15.0	13.5	1.10
10.1330	195	71	15.0	13.0	1.00
10.1430	198	71	15.0	13.0	1.10
10.1530	197	71	15.0	13.2	1.00
10.1630	195	71	15.0	11.8	.90

TABLE J-IX

Raw Data from Run 3

Day:Hour	O_2 in Regenerator Off Gas (% by vol)	SO_2 in Boiler Gas (ppm)	SO_2 in Regenerator Off Gas (μ amp)	CO_2 in Regenerator Off Gas (μ amp)	CO_2 in Boiler Gas (μ amp)
1.0230	5.0	10	10	30	74
1.0330	.0	10	12	38	74
1.0430	.0	20	10	15	84
1.0530	.0	20	10	35	84
1.0630	.0	40	26	27	85
1.0730	1.0	60	14	20	82
1.0830	.0	250	30	36	83
1.0930	.0	325	27	12	83
1.1030	.0	250	10	100	82
1.1130	.0	120	35	39	81
1.1230	.0	80	43	44	82
1.1330	.0	340	34	60	82
1.1430	.0	150	38	56	82
1.1530	.0	220	38	57	84
1.1630	.0	280	46	50	84
1.1730	.0	280	57	37	84
1.1830	.0	280	50	51	84
1.1930	.0	380	49	40	83
1.2030	.0	380	47	40	84
1.2130	.0	150	25	67	85
1.2230	.0	100	46	50	84
1.2330	.0	120	52	38	84
2.0030	.0	0	52	42	84
2.0130	.0	0	49	35	83
2.0230	.0	0	50	39	83
2.0330	.0	0	42	56	83
2.0430	.0	0	38	53	84
2.0530	1.0	0	23	56	82
2.0630	.0	0	35	66	83
2.0730	.0	30	32	94	84
2.0830	13.0	0	16	33	84
2.0930	.0	0	45	68	83
2.1030	.0	0	46	67	84
2.1130	.0	0	42	83	82
2.1230	.0	0	47	60	83
2.1330	.0	8	48	64	82
2.1430	.0	0	29	100	83
2.1530	.0	4	41	85	82
2.1630	.0	0	46	72	82
2.1730	.0	0	45	70	82
2.1830	.0	0	44	69	81

TABLE J-IX

Sheet 17 20

Raw Data from Run 3

Day:Hour	O_2 in Regenerator Off Gas (% by vol)	SO_2 in Boiler Gas (ppm)	SO_2 in Regenerator Off Gas (μ amp)	CO_2 in Regenerator Off Gas (μ amp)	CO_2 in Boiler Gas (μ amp)
2.1930	.0	0	47	60	82
2.2030	.0	4	48	44	82
2.2130	.0	0	48	54	81
2.2230	.0	8	54	46	82
2.2330	.0	3	54	46	81
3.0030	.0	160	52	44	82
3.0130	.0	120	50	52	82
3.0230	.0	120	46	64	82
3.0330	.0	120	46	70	83
3.0430	.0	120	48	52	82
3.0530	.0	11	42	54	82
3.0630	.0	104	48	70	82
3.0730	.0	80	46	74	80
3.0830	.0	64	42	68	80
3.0930	.0	80	39	62	81
3.1030	.0	28	39	62	80
3.1130	.0	20	39	68	80
3.1230	.0	24	40	76	82
3.1330	.0	60	42	79	82
3.1430	.0	0	41	80	77
3.1630	1.0	0	34	34	83
3.1730	.0	0	40	78	81
3.2305	.0	0	40	38	-
4.0005	.0	24	30	46	84
4.0130	.0	4	22	38	85
4.0230	.0	8	28	38	82
4.0330	.0	8	22	40	82
4.0430	.0	0	24	39	85
4.0530	.0	0	28	38	81
4.0630	.0	0	34	39	83
4.0730	.0	0	42	42	84
4.0830	.0	0	38	40	82
4.0930	.0	8	45	40	80
4.1030	.0	8	43	37	81
4.1130	.0	0	32	39	82
4.1230	.0	8	45	37	80
4.1330	.0	0	47	37	80
4.1430	.0	0	46	37	82
4.1530	.0	0	48	37	83
4.1630	.3	0	45	35	83
4.1730	.0	0	45	46	83

TABLE J-IX

Sheet 18 of 20

Raw Data from Run 3

Day:Hour	O_2 in Regenerator Off Gas (% by vol)	SO_2 in Boiler Gas (ppm)	SO_2 in Regenerator Off Gas (μ amp)	CO_2 in Regenerator Off Gas (μ amp)	CO_2 in Boiler Gas (μ amp)
4.1830	.0	0	45	48	83
4.1930	.5	0	44	45	82
5.0130	.0	0	14	60	80
5.0230	.0	16	22	54	82
5.0330	.0	0	38	48	84
5.0430	.0	0	28	38	84
5.0530	.0	0	28	20	82
5.0630	.0	0	40	42	82
5.0730	.0	8	44	41	85
5.0930	.0	32	41	46	86
5.1030	.0	160	44	30	84
5.1130	.0	152	39	44	82
5.1230	.5	88	40	30	81
5.1330	.7	-	39	33	-
5.1430	.3	13	40	38	85
5.1530	.3	32	39	35	84
5.1630	4.5	24	35	15	83
5.1730	17.5	-	40	13	-
5.2200	.0	0	37	23	86
5.2230	1.2	0	24	20	85
5.2330	.2	192	28	48	84
6.0030	.0	112	-	-	86
6.0130	.2	112	38	41	86
6.0230	.3	96	48	42	81
6.0330	4.8	352	36	32	86
6.0430	.4	320	40	46	80
6.0530	.3	320	43	40	86
6.0630	.5	440	43	36	84
6.1130	-	-	-	-	-
6.1230	-	-	-	-	-
6.1330	-	-	-	-	-
6.1430	-	-	-	-	-
6.1500	-	576	-	-	84
6.1830	-	-	-	-	-
6.1930	-	-	-	-	-
6.2030	1.0	224	24	20	85
6.2130	1.0	208	31	24	86
6.2230	.5	200	39	22	85
6.2330	.3	112	45	22	84
7.0030	.5	-	42	26	-
7.0130	.4	-	48	27	-

TABLE J-IX

Sheet 19 of 20

Raw Data from Run 3

Day:Hour	O_2 in Regenerator Off Gas (% by vol)	SO_2 in Boiler Gas (ppm)	SO_2 in Regenerator Off Gas (μ amp)	CO_2 in Regenerator Off Gas (μ amp)	CO_2 in Boiler Gas (μ amp)
7.0230	.4	80	46	27	-
7.0330	.4	152	48	26	84
7.0430	.3	208	49	27	85
7.0530	.4	288	53	23	85
7.0630	.5	160	48	25	86
7.0730	.6	24	47	27	86
7.0830	.1	40	44	30	85
7.0930	.3	12	52	46	85
7.1030	.3	4	49	20	85
7.1130	.3	0	54	34	85
7.1230	.2	200	50	20	86
7.1330	.2	128	49	50	86
7.1430	.2	64	47	52	85
7.1530	.2	32	52	51	84
7.1630	2.0	-	16	52	-
7.1730	.2	376	47	55	86
7.1830	.3	-	43	55	-
7.1930	.3	368	45	55	86
7.2030	.5	360	36	57	87
7.2130	.4	256	43	57	86
7.2230	.4	160	35	57	87
7.2330	.2	176	43	27	85
8.0030	.3	208	36	24	86
8.0130	.2	128	45	29	86
8.0230	.2	64	42	20	80
8.0330	.6	0	37	27	85
8.0430	1.5	16	38	38	86
8.0530	.0	60	45	40	86
8.0630	.0	48	48	44	86
8.0730	.0	8	46	38	85
8.0830	.0	0	48	24	86
8.0930	.0	0	50	26	86
8.1130	.0	8	50	23	84
8.1230	.0	64	50	30	84
8.1330	.0	24	12	14	74
8.1430	.0	32	50	32	84
8.2030	2.0	800	10	14	80
8.2130	.0	160	46	14	84
8.2230	.0	224	46	18	84
8.2330	.4	112	38	21	86
9.0030	.2	72	46	28	85

TABLE J-IX

Sheet 20 of 20

Raw Data from Run 3

Day:Hour	O_2 in Regenerator Off Gas (% by vol)	SO_2 in Boiler Gas (ppm)	SO_2 in Regenerator Off Gas (μ amp)	CO_2 in Regenerator Off Gas (μ amp)	CO_2 in Boiler Gas (μ amp)
9.0130	.2	160	45	20	85
9.0230	.2	192	49	22	86
9.0330	.0	120	45	40	86
9.0430	.3	136	38	26	86
9.0530	.5	96	36	26	86
9.0630	.3	144	40	40	84
9.0730	.6	192	38	35	86
9.0830	.4	176	40	28	86
9.0930	.4	152	39	32	85
9.1030	-	-	-	-	-
9.1130	.4	168	37	39	85
9.1530	15.0	-	12	13	-
9.1630	1.0	-	39	12	-
9.1730	.2	24	41	25	84
9.1830	.3	184	42	19	86
9.1930	.3	232	42	26	86
9.2030	.3	88	42	32	87
9.2130	.0	80	46	34	88
9.2230	.0	32	48	30	86
9.2330	.0	8	48	36	86
10.0030	.0	32	48	34	86
10.0130	.0	48	52	28	87
10.0230	.0	64	50	22	86
10.0330	.0	48	48	22	86
10.0430	-	104	-	-	86
10.0530	.0	48	50	24	87
10.0630	.0	80	48	24	86
10.0730	.0	24	48	26	84
10.0830	.0	32	48	24	86
10.0930	.3	64	44	26	86
10.1030	.3	64	42	25	86
10.1130	.3	96	41	26	87
10.1230	.3	48	44	34	88
10.1330	.2	48	40	28	88
10.1430	.2	44	42	33	88
10.1530	.2	24	46	25	88
10.1630	.4	136	41	28	88

TABLE J-X

CAFB Gas Compositions and Efficiencies

Run 3

TIME DAY.HOUR	BOILER GAS COMP O2 VOL PCT	CO2 VOL PCT	SO2 PPM	REGENERATOR GAS COMP O2 VOL PCT	CO2 VOL PCT	SO2 VOL PCT	SULPHUR REMOVAL PERCENT	REGEN CAO/CAS PERCENT
1.0230	6.8	11.7	10	5.0	1.7	.0	99.2	.2
1.0330	6.8	11.7	10	.0	2.3	.6	99.2	5.5
1.0430	3.2	14.4	20	.0	.5	.0	98.7	.1
1.0530	3.0	14.4	20	.0	2.1	.0	98.7	.1
1.0630	2.8	14.7	40	.0	1.5	3.9	97.4	32.7
1.0730	3.8	13.8	60	1.0	1.0	1.2	95.8	10.4
1.0830	3.7	14.1	250	.0	2.2	4.7	83.0	41.0
1.0930	3.5	14.1	325	.0	.2	4.1	77.9	32.4
1.1030	3.3	13.8	250	.0	10.0	.0	82.6	.2
1.1130	3.3	13.5	120	.0	2.4	5.7	91.5	49.0
1.1230	3.8	13.8	80	.0	2.8	7.2	94.4	62.7
1.1330	3.3	13.8	340	.0	4.3	5.5	76.4	53.5
1.1430	3.2	13.8	150	.0	3.9	6.2	89.6	5[illegible].6
1.1530	3.2	14.4	220	.0	4.0	6.2	85.3	58.1
1.1630	3.0	14.4	280	.0	3.3	7.8	81.3	68.6
1.1730	2.8	14.4	280	.0	2.2	10.0	81.3	80.9
1.1830	3.0	14.4	280	.0	3.4	8.6	81.3	74.9
1.1930	3.0	14.1	380	.0	2.5	8.4	74.1	69.2
1.2030	3.0	14.4	380	.0	2.5	8.0	74.6	66.2
1.2130	2.9	14.7	150	.0	5.1	3.7	90.2	37.7
1.2230	3.1	14.4	100	.0	3.3	7.8	93.3	68.0
1.2330	3.0	14.4	120	.0	2.3	9.0	92.0	73.5
2.0030	3.1	14.4	0	.0	2.6	9.0	100.0	75.5
2.0130	3.2	14.1	0	.0	2.1	8.4	100.0	68.9
2.0230	3.2	14.1	0	.0	2.4	8.6	100.0	71.4
2.0330	3.0	14.1	0	.0	3.9	7.0	100.0	64.4
2.0430	3.0	14.4	0	.0	3.6	6.2	100.0	57.0
2.0530	3.0	13.8	0	1.0	3.9	3.3	100.0	33.7
2.0630	3.0	14.1	0	.0	5.0	5.7	100.0	56.4
2.0730	3.0	14.4	30	.0	8.8	5.1	98.0	67.4
2.0830	3.0	14.4	0	13.0	1.9	1.7	100.0	52.2
2.0930	3.0	14.1	0	.0	5.2	7.6	100.0	76.0
2.1030	3.1	14.4	0	.0	5.1	7.8	100.0	75.6
2.1130	3.5	13.8	0	.0	7.1	7.0	100.0	78.3
2.1230	3.1	14.1	0	.0	4.3	8.0	100.0	73.7
2.1330	3.5	13.8	8	.0	4.7	8.2	99.4	76.0
2.1430	3.5	14.1	0	.0	10.0	4.5	100.0	64.8
2.1530	3.2	13.8	4	.0	7.4	6.8	99.7	77.7
2.1630	3.5	13.8	0	.0	5.7	7.8	100.0	76.9
2.1730	3.2	13.8	0	.0	5.4	7.6	100.0	74.7
2.1830	3.6	13.5	0	.0	5.3	7.4	100.0	71.8

Sheet 2 of 5

TABLE J-X

CAFB Gas Compositions and Efficiencies

Run 3

TIME	BOILER GAS COMP			REGENERATOR GAS COMP				
DAY.HOUR	O2 VOL PCT	CO2 VOL PCT	SO2 PPM	O2 VOL PCT	CO2 VOL PCT	SO2 VOL PCT	SULPHUR REMOVAL PERCENT	REGEN CAO/CAS PERCENT
2.1930	3.2	13.8	0	.0	4.3	8.0	100.0	73.5
2.2030	3.4	13.8	4	.0	2.8	8.2	99.7	69.2
2.2130	3.2	13.5	0	.0	3.7	8.2	100.0	72.5
2.2230	3.5	13.8	8	.0	3.0	9.4	99.4	79.8
2.2330	3.5	13.5	3	.0	3.0	9.4	99.8	78.9
3.0030	3.0	13.8	160	.0	2.8	9.0	88.9	75.8
3.0130	3.2	13.8	120	.0	3.5	8.6	91.7	75.0
3.0230	3.5	13.8	120	.0	4.7	7.8	91.7	73.5
3.0330	3.0	14.1	120	.0	5.4	7.8	91.8	76.4
3.0430	3.5	13.8	120	.0	3.5	8.2	91.7	71.5
3.0530	3.8	13.8	11	.0	3.7	7.0	99.2	62.7
3.0630	3.8	13.8	104	.0	5.4	8.2	92.8	80.0
3.0730	4.0	13.2	80	.0	5.9	7.8	94.2	79.4
3.0830	3.5	13.2	64	.0	5.2	7.0	95.4	68.2
3.0930	8.5	13.5	80	.0	4.5	6.4	94.3	60.6
3.1030	4.0	13.2	28	.0	4.5	6.4	98.0	60.5
3.1130	4.0	13.2	20	.0	5.2	6.4	98.6	63.4
3.1230	3.5	13.8	24	.0	6.2	6.6	98.3	69.1
3.1330	3.5	13.8	60	.0	6.5	7.0	95.8	74.3
3.1430	5.8	12.4	0	.0	6.7	6.8	100.0	73.4
3.1630	3.4	14.1	0	1.0	2.0	5.5	100.0	48.1
3.1730	3.5	13.5	0	.0	6.4	6.6	100.0	71.6
3.2305	-	-	-	.0	2.3	6.6	-	54.4
4.0005	3.5	14.4	24	.0	3.0	4.7	98.4	40.7
4.0130	2.5	14.7	4	.0	2.3	3.1	99.7	26.4
4.0230	2.8	13.8	8	.0	2.3	4.3	99.4	36.3
4.0330	2.5	13.8	8	.0	2.5	3.1	99.4	26.5
4.0430	2.5	14.7	0	.0	2.4	3.5	100.0	30.1
4.0530	3.0	13.5	0	.0	2.3	4.3	100.0	36.1
4.0630	2.5	14.1	0	.0	2.4	5.5	100.0	45.4
4.0730	2.2	14.4	0	.0	2.6	7.0	100.0	58.1
4.0830	3.1	13.8	0	.0	2.5	6.2	100.0	51.9
4.0930	3.6	13.2	8	.0	2.5	7.6	99.4	61.8
4.1030	3.6	13.5	8	.0	2.2	7.2	99.4	59.1
4.1130	3.1	13.8	0	.0	2.4	5.1	100.0	42.8
4.1230	3.5	13.2	8	.0	2.2	7.6	99.4	61.5
4.1330	3.5	13.2	0	.0	2.2	8.0	100.0	64.6
4.1430	3.0	13.8	0	.0	2.2	7.8	100.0	63.2
4.1530	3.0	14.1	0	.0	2.2	8.2	100.0	66.0
4.1630	3.0	14.1	0	.3	2.1	7.6	100.0	62.2
4.1730	3.0	14.1	0	.0	3.0	7.6	100.0	63.8

Sheet 3 of 5

TABLE J-X

CAFB Gas Compositions and Efficiencies

Run 3

TIME DAY.HOUR	BOILER GAS COMP O2 VOL PCT	 CO2 VOL PCT	 SO2 PPM	REGENERATOR GAS COMP O2 VOL PCT	 CO2 VOL PCT	 SO2 VOL PCT	 SULPHUR REMOVAL PERCENT	 REGEN CAO/CAS PERCENT
4.1830	3.0	14.1	0	.0	3.1	7.6	100.0	64.5
4.1930	3.0	13.8	0	.5	2.9	7.4	100.0	64.2
5.0130	2.0	13.2	0	.0	4.3	1.2	100.0	-
5.0230	3.0	13.8	16	.0	3.7	3.1	98.9	-
5.0330	2.5	14.4	0	.0	3.1	6.2	100.0	54.0
5.0430	3.0	14.4	0	.0	2.3	4.3	100.0	36.3
5.0530	3.0	13.8	0	.0	1.0	4.3	100.0	37.9
5.0630	3.0	13.8	0	.0	2.6	6.6	100.0	55.3
5.0730	2.8	14.7	8	.0	2.5	7.4	99.5	61.2
5.0930	2.0	15.0	32	.0	3.0	6.8	97.9	58.2
5.1030	3.0	14.4	160	.0	1.7	7.4	89.3	58.5
5.1130	3.0	13.8	152	.0	2.8	6.4	89.4	54.5
5.1230	3.5	13.5	88	.5	1.7	6.6	93.7	54.1
5.1330	-	-	-	.7	1.9	6.4	-	53.8
5.1430	2.8	14.7	13	.3	2.3	6.6	99.1	55.2
5.1530	2.9	14.4	32	.3	2.1	6.4	97.9	53.3
5.1630	3.0	14.1	24	4.5	.5	5.7	98.4	55.0
5.1730	3.1	-	-	17.5	.3	6.6	-	-
5.2200	4.2	15.0	0	.0	1.2	6.0	100.0	-
5.2230	2.8	14.7	0	1.2	1.0	3.5	100.0	29.6
5.2330	4.7	14.4	192	.2	3.1	4.3	87.2	38.5
6.0030	5.7	15.0	112	.0	-	-	92.8	-
6.0130	2.0	15.0	112	.2	2.5	6.2	92.8	52.8
6.0230	3.4	13.5	96	.3	2.6	7.2	93.2	68.8
6.0330	1.8	15.0	352	4.8	1.9	5.8	77.4	62.6
6.0430	3.0	13.2	320	.4	3.0	6.6	76.8	58.0
6.0530	2.0	15.0	320	.3	2.5	7.2	79.4	60.3
6.0630	1.6	14.4	440	.5	2.2	7.2	70.6	59.9
6.1130	-	-	-	-	-	-	-	-
6.1230	-	-	-	-	-	-	-	
6.1330	-	-	-	-	-	-	-	-
6.1430	-	-	-	-	-	-	-	-
6.1500	2.4	14.4	576	-	-	-	61.6	-
6.1830	-	-	-	-	-	-	-	-
6.1930	-	-	-	-	-	-	-	-
6.2030	2.4	14.7	224	1.0	1.0	3.5	85.3	29.5
6.2130	2.0	15.0	208	1.0	1.3	4.9	86.7	41.4
6.2230	2.0	14.7	200	.5	1.1	6.4	86.9	51.7
6.2330	2.0	14.4	112	.3	1.1	7.6	92.5	59.8
7.0030	2.0	-	96	.5	1.4	7.0	-	56.8
7.0130	2.1	-	16	.4	1.5	8.2	-	65.9

TABLE J-X

CAFB Gas Compositions and Efficiencies

Run 3

TIME	BOILER GAS COMP			REGENERATOR GAS COMP				
DAY.HOUR	O2 VOL PCT	CO2 VOL PCT	SO2 PPM	O2 VOL PCT	CO2 VOL PCT	SO2 VOL PCT	SULPHUR REMOVAL PERCENT	REGEN CAO/CAS PERCENT
7.0230	2.0	-	80	.4	1.5	7.8	-	62.8
7.0330	2.2	14.4	152	.4	1.4	8.2	89.9	65.6
7.0430	2.0	14.7	208	.3	1.5	8.4	86.4	67.1
7.0530	1.8	14.7	288	.4	1.2	9.2	81.2	72.3
7.0630	2.0	15.0	160	.5	1.3	8.2	89.7	65.7
7.0730	2.0	15.0	24	.6	1.5	8.0	98.5	65.0
7.0830	2.2	14.7	40	.1	1.7	7.4	97.4	59.7
7.0930	2.4	14.7	12	.3	3.0	9.0	99.2	77.1
7.1030	2.3	14.7	4	.3	1.0	8.4	99.7	65.4
7.1130	2.0	14.7	0	.3	2.0	9.4	100.0	76.7
7.1230	2.0	15.0	200	.2	1.0	8.6	87.2	66.6
7.1330	2.3	15.0	128	.2	3.3	8.4	91.8	82.9
7.1430	2.1	14.7	64	.2	3.5	8.0	95.8	79.1
7.1530	2.8	14.4	32	.2	3.4	9.0	97.9	87.4
7.1630	-	-	16	2.0	3.5	1.7	-	20.8
7.1730	2.0	15.0	376	.2	3.8	8.0	75.9	73.8
7.1830	-	-	-	.3	3.8	7.2	-	67.7
7.1930	2.2	15.0	368	.3	3.8	7.6	76.4	71.0
7.2030	1.8	15.3	360	.5	4.0	5.8	77.4	57.3
7.2130	2.0	15.0	256	.4	4.0	7.2	83.6	69.0
7.2230	2.0	15.3	160	.4	4.0	5.7	90.0	55.4
7.2330	2.0	14.7	176	.2	1.5	7.2	88.5	59.5
8.0030	2.2	15.0	208	.3	1.3	5.8	86.7	48.8
8.0130	2.0	15.0	128	.2	1.6	7.6	91.8	63.0
8.0230	2.7	13.2	64	.2	1.0	7.0	95.4	56.5
8.0330	2.0	14.7	0	.6	1.5	6.0	100.0	51.3
8.0430	2.0	15.0	16	1.5	2.3	6.2	99.3	58.3
8.0530	2.0	15.0	60	.0	2.5	7.6	96.[illegible]	65.1
8.0630	2.0	15.0	48	.0	2.8	8.2	96.9	71.0
8.0730	2.1	14.7	8	.0	2.3	7.8	99.5	66.7
8.0830	1.9	15.0	0	.0	1.3	8.2	100.0	66.2
8.0930	2.1	15.0	0	.0	1.4	8.6	100.0	69.6
8.1130	3.0	14.4	8	.0	1.2	8.6	99.5	68.7
8.1230	2.8	14.4	64	.0	1.7	8.6	95.7	70.5
8.1330	6.0	11.7	24	.0	.4	.6	98.0	5.1
8.1430	3.6	14.4	32	.0	1.9	8.6	97.8	70.6
8.2030	5.2	13.2	800	2.0	.4	.0	41.9	-
8.2130	3.0	14.4	160	.0	.4	7.8	89.3	60.4
8.2230	3.1	14.4	224	.0	.8	7.8	85.0	61.7
8.2330	2.5	15.0	112	.4	1.0	6.2	92.8	51.9
9.0030	2.5	14.7	72	.2	1.6	7.8	95.3	64.8

TABLE J-X

CAFB Gas Compositions and Efficiencies

Run 3

TIME	BOILER GAS COMP			REGENERATOR GAS COMP				
DAY.HOUR	O2 VOL PCT	CO2 VOL PCT	SO2 PPM	O2 VOL PCT	CO2 VOL PCT	SO2 VOL PCT	SULPHUR REMOVAL PERCENT	REGEN CAO/CAS PERCENT
9.0130	2.7	14.7	160	.2	1.0	7.6	89.5	61.3
9.0230	2.2	15.0	192	.2	1.1	8.4	87.7	62.3
9.0330	2.3	15.0	120	.0	2.5	7.6	92.3	65.
9.0430	2.2	15.0	136	.3	1.4	6.2	91.3	51.2
9.0530	2.0	15.0	96	.5	1.4	5.8	93.8	48.6
9.0630	2.8	14.4	144	.3	2.5	6.6	90.4	57.4
9.0730	2.0	15.0	192	.6	2.1	6.2	87.7	53.9
9.0830	2.5	15.0	176	.4	1.6	6.6	88.7	54.9
9.0930	2.8	14.7	152	.4	1.9	6.4	90.0	54.2
9.1030	-	-	-	3.5	2.5	-	77.7	-
9.1130	2.5	14.7	168	.4	2.4	6.0	89.0	52.8
9.1530	-	-	-	15.0	.3	.6	-	-
9.1630	-	-	-	1.0	.2	6.4	-	53.7
9.1730	2.6	14.5	24	.2	1.3	6.8	98.4	57.7
9.1830	2.1	15.0	184	.3	.9	7.0	88.2	57.8
9.1930	2.1	15.0	232	.3	1.4	7.0	85.1	59.3
9.2030	1.8	15.3	88	.3	1.9	7.0	94.5	61.1
9.2130	1.2	15.6	80	.0	2.0	7.8	95.1	67.4
9.2230	1.8	15.0	32	.0	1.7	8.2	97.9	69.5
9.2330	2.0	15.0	8	.0	2.2	8.2	99.5	71.5
10.0030	2.0	15.0	32	.0	2.0	8.2	97.9	71.2
10.0130	1.8	15.3	48	.0	1.6	9.0	97.0	75.6
10.0230	2.0	15.0	64	.0	1.1	8.6	95.9	70.9
10.0330	1.9	15.0	48	.0	1.1	8.2	96.9	67.4
10.0430	1.8	15.0	104	-	-	-	93.3	-
10.0530	1.6	15.3	48	.0	1.3	8.6	97.0	71.0
10.0630	1.7	15.0	80	.0	1.3	8.2	94.9	69.4
10.0730	2.2	14.4	24	.0	1.4	8.2	98.4	69.8
10.0830	2.0	15.0	32	.0	1.3	8.2	97.9	69.3
10.0930	1.6	15.0	64	.3	1.4	7.4	95.9	64.6
10.1030	1.6	15.0	64	.3	1.3	7.0	95.9	59.1
10.1130	1.3	15.3	96	.3	1.4	6.8	93.9	57.8
10.1230	1.0	15.6	48	.3	2.0	7.4	97.0	64.4
10.1330	1.0	15.6	48	.2	1.6	6.6	97.0	56.5
10.1430	1.0	15.6	44	.2	1.9	7.0	97.3	60.7
10.1530	1.0	15.6	24	.2	1.3	7.8	98.5	64.9
10.1630	.9	15.6	136	.4	1.6	6.8	91.6	58.7

TABLE J-XI

Sheet 1 of 5

CAFB Feed Ratios and Regenerator Sulphur

Run 3

DAY.HOUR	VELOCITY FT/SEC	AIR/FUEL PCT STOIC	CA/S RATIO MOL/MOL	REGENERATOR SULPHUR LB/HR	PCT OF FED
1.0230	3.39	24.2	1.35	.01	.2
1.0330	3.34	24.0	1.35	.57	7.6
1.0430	3.91	25.3	1.21	.01	.1
1.0530	3.89	21.8	1.08	.01	.1
1.0630	3.67	20.8	1.08	3.35	35.6
1.0730	3.65	20.9	1.08	.91	9.6
1.0830	3.64	20.9	1.11	3.36	35.7
1.0930	3.66	20.9	1.11	3.09	32.8
1.1030	3.66	20.8	1.11	.01	.1
1.1130	3.79	20.9	1.11	4.47	47.5
1.1230	3.92	22.0	1.14	5.80	61.6
1.1330	3.72	20.0	1.11	4.03	42.8
1.1430	3.93	21.4	1.11	4.82	51.2
1.1530	3.78	22.3	1.53	5.39	57.2
1.1630	3.89	22.4	1.53	6.80	72.2
1.1730	3.76	21.8	1.53	9.86	104.7
1.1830	3.76	21.8	1.53	8.31	88.2
1.1930	3.74	21.0	1.53	9.05	96.1
1.2030	3.74	21.4	1.93	8.61	91.4
1.2130	3.66	21.3	1.93	3.61	38.3
1.2230	3.61	21.4	1.93	7.76	82.4
1.2330	3.58	21.0	1.93	9.02	95.8
2.0030	3.60	21.5	2.58	9.21	97.8
2.0130	3.85	22.5	2.58	8.33	88.4
2.0230	3.66	22.0	2.58	8.76	93.1
2.0330	3.67	22.0	3.40	7.14	75.8
2.0430	3.68	22.0	3.40	6.19	65.7
2.0530	3.62	21.7	2.15	3.25	34.5
2.0630	3.60	21.7	2.15	5.76	61.1
2.0730	3.62	21.7	2.15	5.31	56.4
2.0830	3.63	21.7	2.48	1.69	18.0
2.0931	3.65	21.7	2.48	6.63	70.4
2.1030	3.66	21.7	2.48	8.82	93.7
2.1130	3.62	21.8	1.93	8.05	85.5
2.1230	3.64	21.7	1.91	6.23	66.1
2.1330	3.66	21.8	1.93	8.82	93.7
2.1430	3.65	21.8	1.93	4.38	46.6
2.1530	3.67	21.7	1.47	5.38	57.1
2.1630	3.67	21.8	1.47	8.17	86.7
2.1730	3.61	21.3	1.47	6.44	68.3
2.1830	3.53	20.9	1.47	7.29	77.4

TABLE J-XI

Sheet 2 of 5

CAFB Feed Ratios and Regenerator Sulphur

Run 3

DAY.HOUR	VELOCITY FT/SEC	AIR/FUEL PCT STOIC	CA/S RATIO MOL/MOL	REGENERATOR SULPHUR LB/HR	PCT OF FED
2.1930	3.55	20.8	1.47	6.30	66.9
2.2030	3.58	20.9	1.47	6.29	66.8
2.2130	3.52	20.4	1.08	6.75	71.6
2.2230	3.68	21.3	1.08	6.52	69.2
2.2330	3.52	20.4	1.08	7.92	84.1
3.0030	3.36	19.9	1.08	6.31	67.0
3.0130	3.46	20.4	1.08	6.92	73.5
3.0230	3.55	21.3	1.08	6.95	73.8
3.0330	3.45	20.3	1.08	7.28	77.3
3.0430	3.47	20.4	1.08	7.60	80.7
3.0530	3.41	20.5	1.53	6.43	68.3
3.0630	3.42	20.5	1.53	7.62	80.9
3.0730	3.42	20.5	1.53	6.49	68.9
3.0830	3.29	19.5	1.53	6.78	72.0
3.0930	3.21	20.1	1.53	6.14	65.2
3.1030	3.28	19.6	1.73	6.30	66.9
3.1130	3.35	20.1	1.73	5.69	60.4
3.1230	3.10	19.2	1.73	6.11	64.9
3.1330	3.12	19.2	1.73	6.96	74.0
3.1430	3.17	19.6	1.01	6.28	66.7
3.1630	3.37	19.9	1.01	4.52	48.0
3.1730	3.41	20.0	1.01	4.72	50.1
3.2305	3.52	19.4	1.02	5.94	63.4
4.0005	3.73	22.1	1.02	4.60	49.1
4.0130	3.88	22.1	3.58	2.44	26.1
4.0230	3.98	23.1	3.58	3.74	39.9
4.0330	3.97	23.0	.79	2.57	27.4
4.0430	3.89	22.9	.79	2.56	27.3
4.0530	3.99	23.0	.79	4.45	47.5
4.0630	3.85	21.8	.79	5.74	61.2
4.0730	3.73	21.3	.79	6.77	72.2
4.0830	3.85	21.9	.79	5.38	57.4
4.0930	3.95	22.9	.79	8.54	91.2
4.1030	3.87	22.9	.79	5.22	55.7
4.1130	3.75	21.9	1.51	4.08	43.5
4.1230	3.81	21.9	1.51	6.89	73.6
4.1330	3.73	21.4	1.51	6.74	71.9
4.1430	3.73	21.3	1.51	6.49	69.2
4.1530	3.69	21.0	1.51	7.20	76.8
4.1630	3.61	20.9	1.34	6.25	66.7
4.1730	3.62	20.7	1.34	7.28	77.7

TABLE J-XI

Sheet 3 of 5

CAFB Feed Ratios and Regenerator Sulphur

Run 3

DAY.HOUR	VELOCITY FT/SEC	AIR/FUEL PCT STOIC	CA/S RATIO MOL/MOL	REGENERATOR SULPHUR LB/HR	PCT OF FED
4.1830	3.54	20.3	1.34	7.11	75.9
4.1930	3.54	20.7	1.34	5.84	62.3
5.0130	3.87	22.8	1.75	.04	.4
5.0230	3.79	21.7	1.69	.10	1.1
5.0330	3.70	21.4	1.73	5.12	56.4
5.0430	3.66	22.0	2.22	3.72	42.1
5.0530	3.64	21.5	2.17	3.74	41.2
5.0630	3.64	21.5	2.17	5.99	66.0
5.0730	3.72	22.0	2.22	6.68	75.4
5.0930	3.52	20.6	1.27	5.32	59.5
5.1030	3.75	22.6	1.27	6.43	71.9
5.1130	3.64	21.7	1.27	5.18	57.9
5.1230	3.61	21.4	1.27	5.94	66.4
5.1330	3.50	19.7	1.27	5.65	63.2
5.1430	3.42	20.0	1.27	5.89	65.9
5.1530	3.37	19.6	1.27	5.09	56.9
5.1630	3.10	16.9	1.99	4.60	51.5
5.1730	2.78	15.4	1.99	6.42	71.8
5.2200	3.73	21.3	2.00	.21	2.4
5.2230	3.63	19.7	2.00	3.04	34.2
5.2330	3.37	21.6	2.00	3.86	43.4
6.0030	3.35	21.4	.55	-	-
6.0130	3.10	19.5	.55	5.36	60.2
6.0230	2.99	18.0	.55	6.82	76.7
6.0330	3.13	18.2	.55	5.95	66.9
6.0430	3.06	18.5	.55	5.29	59.4
6.0530	3.08	18.2	.55	6.74	75.8
6.0630	3.07	17.7	.55	6.67	75.0
6.1130	3.90	19.4	.52	-	-
6.1230	-	20.2	.51	-	-
6.1330	-	19.9	.51	-	-
6.1430	-	15.9	.51	-	-
6.1500	3.27	18.3	.52	-	-
6.1830	-	23.3	1.80	-	-
6.1930	-	21.4	1.80	-	-
6.2030	3.79	19.7	1.81	2.97	31.2
6.2130	3.82	19.6	1.81	4.02	42.2
6.2230	3.93	20.7	1.80	5.23	54.6
6.2330	3.99	20.7	1.35	6.31	65.9
7.0030	3.95	21.2	1.35	5.91	61.7
7.0130	3.95	21.2	1.35	6.37	66.5

TABLE J-XI

Sheet 4 of

CAFB Feed Ratios and Regenerator Sulphur

Run 3

DAY.HOUR	VELOCITY FT/SEC	AIR/FUEL PCT STOIC	CA/S RATIO MOL/MOL	REGENERATOR SULPHUR LB/HR	PCT OF FED
7.0230	3.98	21.2	1.35	6.21	64.8
7.0330	4.00	21.2	1.35	6.38	66.6
7.0430	3.98	21.2	1.35	6.40	66.8
7.0530	4.01	21.1	1.89	7.02	73.3
7.0630	3.94	21.2	1.89	6.37	66.5
7.0730	3.89	20.7	1.89	6.29	65.7
7.0830	3.84	20.8	1.89	5.81	60.7
7.0930	3.74	20.3	1.16	7.18	75.0
7.1030	3.79	20.3	1.16	6.16	64.4
7.1130	3.82	20.3	1.16	6.92	72.2
7.1230	3.85	20.3	1.16	6.17	64.4
7.1330	3.86	20.4	1.16	7.65	79.9
7.1430	4.38	20.3	1.16	7.91	82.6
7.1530	3.86	20.2	.90	8.95	93.5
7.1630	3.86	19.1	.90	1.61	16.8
7.1730	3.80	19.4	.90	7.45	77.8
7.1830	-	18.7	.90	6.56	68.5
7.1930	3.84	19.5	.90	6.95	72.6
7.2030	3.85	19.3	.90	5.24	54.7
7.2130	3.71	18.5	.90	6.48	67.7
7.2230	3.69	18.5	.90	4.90	51.2
7.2330	3.70	18.5	1.51	6.07	63.4
8.0030	3.73	19.0	1.51	4.77	49.8
8.0130	3.72	18.9	1.51	6.31	65.9
8.0230	3.68	18.9	1.51	5.86	61.2
8.0330	3.76	19.5	1.51	5.52	57.6
8.0430	3.77	19.6	2.05	5.61	58.6
8.0530	3.73	19.5	2.05	6.47	67.6
8.0630	3.75	19.5	2.05	7.08	73.9
8.0730	3.69	19.5	.90	5.93	61.9
8.0830	3.76	24.9	1.15	6.12	81.7
8.0930	3.79	25.0	1.15	6.60	88.1
8.1130	3.70	25.3	1.15	6.75	90.1
8.1230	3.70	25.3	1.15	6.81	90.9
8.1330	3.76	26.7	1.15	.44	5.9
8.1430	3.67	25.7	1.64	7.32	97.8
8.2030	4.30	22.2	1.39	.00	.0
8.2130	4.14	22.7	1.39	6.27	71.1
8.2230	4.27	22.9	1.39	6.03	68.3
8.2330	3.83	20.5	1.39	4.76	53.9
9.0030	3.98	20.6	1.81	6.05	68.6

TABLE J-XI

Sheet 5 of 5

CAFB Feed Ratios and Regenerator Sulphur

Run 3

DAY.HOUR	VELOCITY FT/SEC	AIR/FUEL PCT STOIC	CA/S RATIO MOL/MOL	REGENERATOR SULPHUR	
				LB/HR	PC\ OF FED
9.0130	3.99	22.4	1.81	5.95	67.4
9.0230	3.94	21.7	1.81	5.93	67.2
9.0330	4.12	22.3	1.81	5.52	62.5
9.0430	4.05	21.8	1.81	5.74	65.0
9.0530	4.11	22.2	1.04	5.60	63.4
9.0630	3.93	21.1	1.04	5.98	67.7
9.0730	4.06	20.9	1.04	5.76	65.2
9.0830	4.19	21.7	1.04	6.06	68.7
9.0930	4.09	21.7	1.04	5.87	66.5
9.1030	4.27	-	-	-	-
9.1130	3.95	21.0	1.04	5.43	61.5
9.1530	3.74	20.4	1.05	.07	.8
9.1630	-	21.4	.91	5.84	66.7
9.1730	3.91	21.6	.91	6.18	70.6
9.1830	3.88	21.4	.91	6.61	75.5
9.1930	3.91	21.8	.91	6.88	78.6
9.2030	3.78	21.2	.91	6.52	74.4
9.2130	3.94	22.0	.91	6.69	76.3
9.2230	3.99	22.2	.91	6.99	79.7
9.2330	3.74	20.4	.91	6.80	77.7
10.0030	3.89	21.4	.91	6.58	75.1
10.0130	3.88	21.3	.98	7.34	83.8
10.0230	3.89	21.4	.98	6.94	79.2
10.0330	3.89	21.3	.98	6.94	79.2
10.0430	-	-	.98	-	-
10.0530	3.88	21.2	.98	7.26	82.9
10.0630	3.92	21.2	.98	6.97	79.5
10.0730	4.03	22.4	.98	7.05	80.5
10.0830	3.91	22.2	.53	6.98	79.7
10.0930	3.68	22.2	.58	6.38	80.3
10.1030	3.88	23.3	.58	5.74	72.3
10.1130	3.80	23.1	.58	5.60	70.5
10.1230	3.77	23.0	.58	6.11	76.9
10.1330	3.72	22.5	.58	5.38	67.8
10.1430	3.72	22.5	.58	5.73	72.1
10.1530	3.73	22.5	.58	6.30	79.3
10.1630	3.28	17.6	.58	5.50	69.2

TABLE J-XII

Sheet 1 of 2

CAFB Pilot Plant Solids Circulation Rates Computed from Heat Balance

Run 3

FLOW OF HOT SOLIDS IN POUNDS PER HOUR

TIME DAY.HOUR	CAO CIRCULATION	TOTAL SOLIDS FLOW TO REGENERATOR	FROM REGENERATOR
1.0230	6321	6336	6350
1.0330	1927	1949	1965
1.0430	1764	1781	1796
1.0530	933	947	959
1.0630	673	695	700
1.0730	577	596	608
1.0830	485	502	504
1.0930	574	594	600
1.1030	861	873	878
1.1130	1015	1035	1034
1.1230	618	639	632
1.1330	648	665	662
1.1430	746	764	759
1.1530	639	660	654
1.1630	564	586	576
1.1730	463	490	473
1.1830	516	541	527
1.1930	687	715	703
1.2030	656	684	673
1.2130	710	732	734
1.2230	469	494	484
1.2330	551	577	564
2.0030	508	535	520
2.0130	640	666	655
2.0230	549	576	563
2.0330	546	571	562
2.0430	571	595	590
2.0530	757	779	782
2.0630	643	666	661
2.0730	597	617	607
2.0830	-	-	-
2.0930	388	405	396
2.1030	532	558	543
2.1130	440	465	449
2.1230	345	364	354
2.1330	498	524	509
2.1430	442	459	451
2.1530	350	366	356
2.1630	498	522	507
2.1730	395	415	404
2.1830	474	497	485

TABLE J-XII

Sheet 2 of 5

CAFB Pilot Plant Solids Circulation Rates Computed from Heat Balance

Run 3

FLOW OF HOT SOLIDS IN POUNDS PER HOUR

TIME DAY.HOUR	CAO CIRCULATION	TOTAL SOLIDS FLOW TO REGENERATOR	FROM REGENERATOR
2.1930	383	402	392
2.2030	430	450	441
2.2130	449	469	459
2.2230	366	384	373
2.2330	438	460	446
3.0030	319	338	327
3.0130	385	406	395
3.0230	414	436	424
3.0330	419	441	428
3.0430	472	495	484
3.0530	472	495	487
3.0630	377	399	384
3.0730	327	345	333
3.0830	449	471	461
3.0930	512	535	528
3.1030	501	524	517
3.1130	431	451	444
3.1230	424	444	435
3.1330	434	456	444
3.1430	402	422	411
3.1630	536	556	555
3.1730	386	402	394
3.2305	592	616	612
4.0005	695	720	722
4.0130	634	654	661
4.0230	1028	1051	1054
4.0330	1095	1116	1123
4.0430	856	874	880
4.0530	1275	1301	1306
4.0630	1078	1105	1105
4.0730	611	637	631
4.0830	749	771	768
4.0930	858	888	879
4.1030	467	486	482
4.1130	828	849	850
4.1230	703	728	720
4.1330	584	607	599
4.1430	562	584	577
4.1530	599	623	614
4.1630	614	635	629
4.1730	664	689	680

TABLE J-XII

Sheet 3 of [illegible]

CAFB Pilot Plant Solids Circulation Rates Computed from Heat Balance

Run 3

FLOW OF HOT SOLIDS IN POUNDS PER HOUR

TIME DAY.HOUR	CAO CIRCULATION	TOTAL SOLIDS FLOW TO REGENERATOR	FROM REGENERATOR
4.1830	573	597	588
4.1930	479	499	4[illegible]2
5.0130	-	-	-
5.0230	-	-	-
5.0330	654	675	671
5.0430	684	706	709
5.0530	743	766	772
5.0630	600	623	619
5.0730	631	654	648
5.0930	497	516	512
5.1030	677	701	695
5.1130	719	740	736
5.1230	684	707	704
5.1330	652	675	671
5.1430	684	707	702
5.1530	585	605	602
5.1630	289	306	304
5.1730	-	-	-
5.2200	-	-	-
5.2230	881	903	910
5.2330	909	931	934
6.0030	1492	1513	1545
6.0130	552	574	571
6.0230	478	500	490
6.0330	413	434	427
6.0430	448	467	463
6.0530	551	576	569
6.0630	541	565	559
6.1130	221	222	221
6.2030	661	682	689
6.2130	518	539	541
6.2230	543	565	563
6.2330	551	573	568
7.0030	523	545	540
7.0130	428	448	441

TABLE J-XII

Sheet 4 of 5

CAFB Pilot Plant Solids Circulation Rates Computed from Heat Balance

Run 3

FLOW OF HOT SOLIDS IN POUNDS PER HOUR

TIME DAY.HOUR	CAO CIRCULATION	TOTAL SOLIDS FLOW TO REGENERATOR	FROM REGENERATOR
7.0230	468	489	483
7.0330	441	462	454
7.0430	422	442	434
7.0530	438	458	448
7.0630	426	447	439
7.0730	433	454	447
7.0830	441	462	457
7.0930	354	374	362
7.1030	408	428	421
7.1130	362	382	371
7.1230	458	478	471
7.1330	336	356	342
7.1430	398	420	406
7.1530	362	384	367
7.1630	514	531	538
7.1730	449	471	459
7.1930	452	474	463
7.2030	458	478	473
7.2130	417	437	428
7.2230	424	444	440
7.2330	449	471	465
8.0030	445	466	465
8.0130	406	427	421
8.0230	481	503	498
8.0330	498	521	519
8.0430	401	422	417
8.0530	420	441	433
8.0630	460	482	472
8.0730	421	440	432
8.0830	427	447	439
8.0930	479	499	490
8.1130	505	526	517
8.1230	462	482	473
8.1330	-	-	-
8.1430	495	518	508
8.2030	146	146	146
8.2130	626	648	642
8.2230	570	591	585
8.2330	514	534	532
9.0030	471	491	484

TABLE J-XII

Sheet 5 [illegible]

CAFB Pilot Plant Solids Circulation Rates Computed from Heat Balance

Run 3

FLOW OF HOT SOLIDS IN POUNDS PER HOUR

TIME DAY.HOUR	CAO CIRCULATION	TOTAL SOLIDS FLOW TO REGENERATOR	TOTAL SOLIDS FLOW FROM REGENERATOR
9.0130	514	534	528
9.0230	536	557	551
9.0330	448	467	460
9.0430	782	806	803
9.0530	734	759	758
9.0630	639	662	657
9.0730	649	672	669
9.0830	652	675	671
9.0930	640	664	660
9.1030	-	-	-
9.1130	590	612	609
9.1530	175	173	171
9.1730	624	647	642
9.1830	610	634	629
9.1930	710	734	728
9.2030	519	542	535
9.2130	406	427	419
9.2230	520	541	532
9.2330	455	476	466
10.0030	435	454	445
10.0130	411	431	420
10.0230	421	442	432
10.0330	459	481	472
10.0430	919	937	965
10.0530	479	501	491
10.0630	511	532	523
10.0730	558	580	571
10.0830	488	510	500
10.0930	488	509	502
10.1030	521	542	537
10.1130	541	562	557
10.1230	457	477	470
10.1330	510	530	526
10.1430	480	500	494
10.1530	476	496	489
10.1630	445	465	461

H. Figures from Continuous Pilot Plant Runs

CAFB Pilot Plant Operation Log (Run 1)

Sulphur Removal Efficiency and Lime Addition Rate

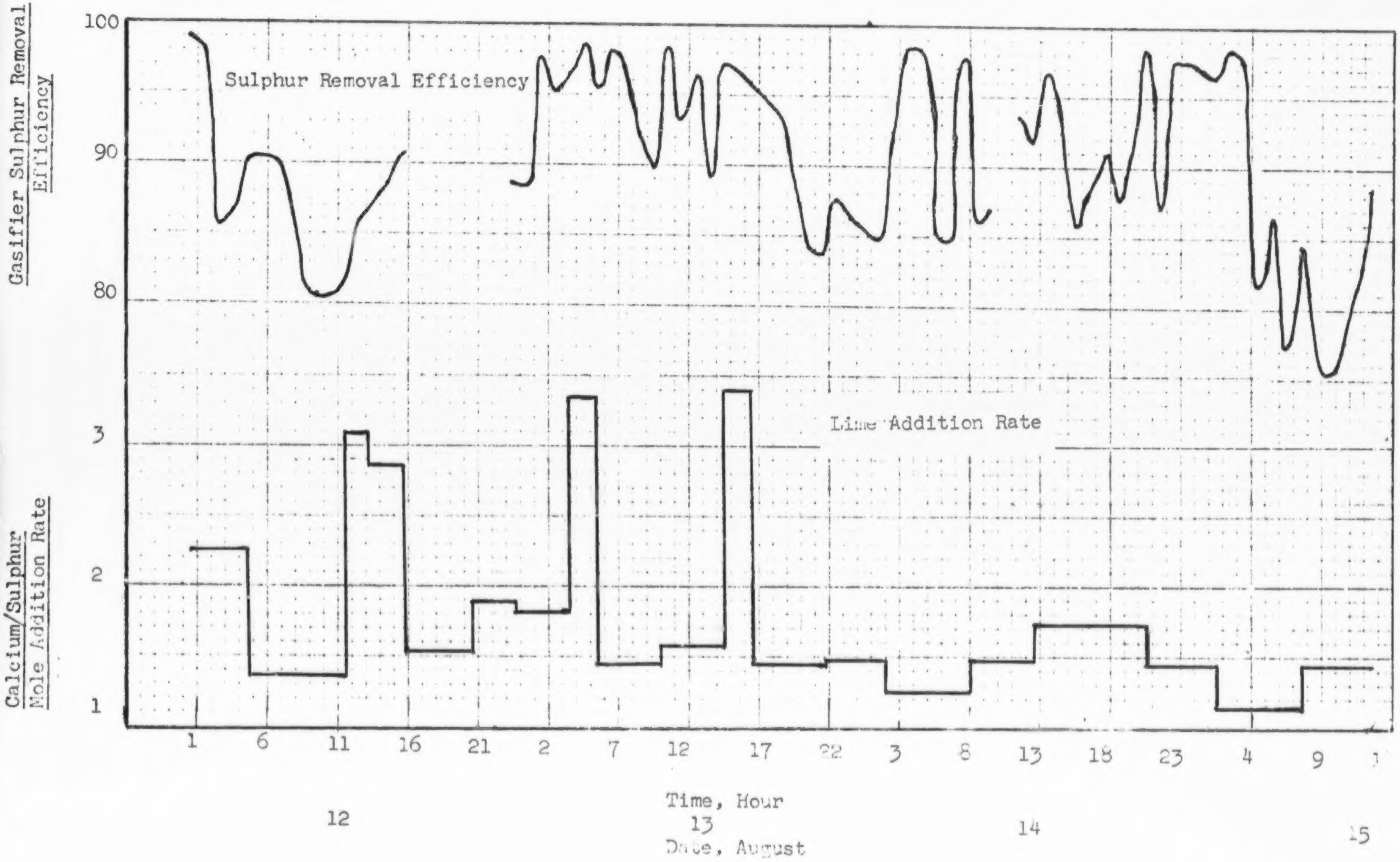


Fig.J-1

CAFB Pilot Plant Operation Log (Run 1)

Air Fuel Ratio and Superficial Gas Velocity

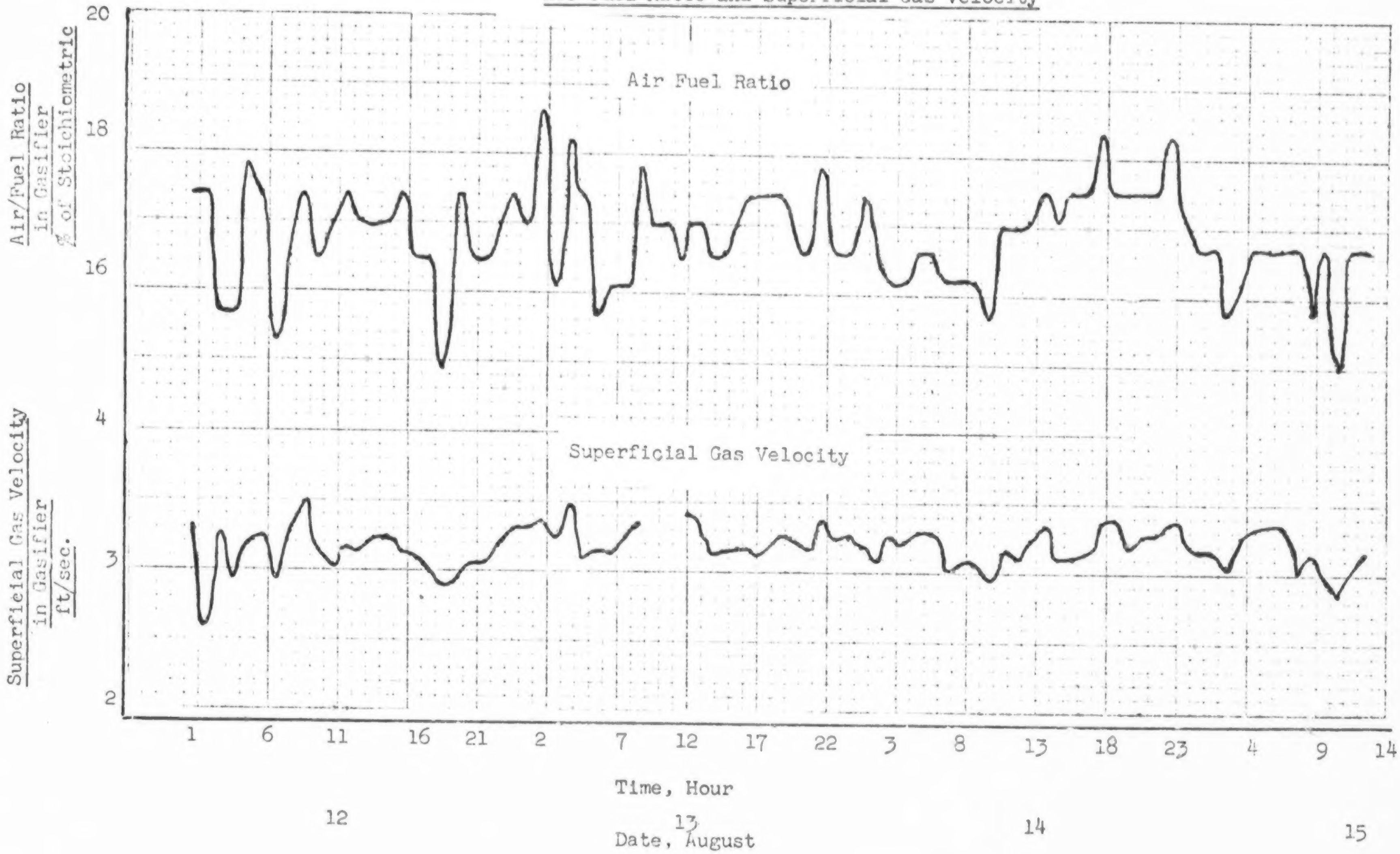


Fig J-2

CAFB Pilot Plant Operation Log (Run 1)

Gasifier Pressure and Bed Depth

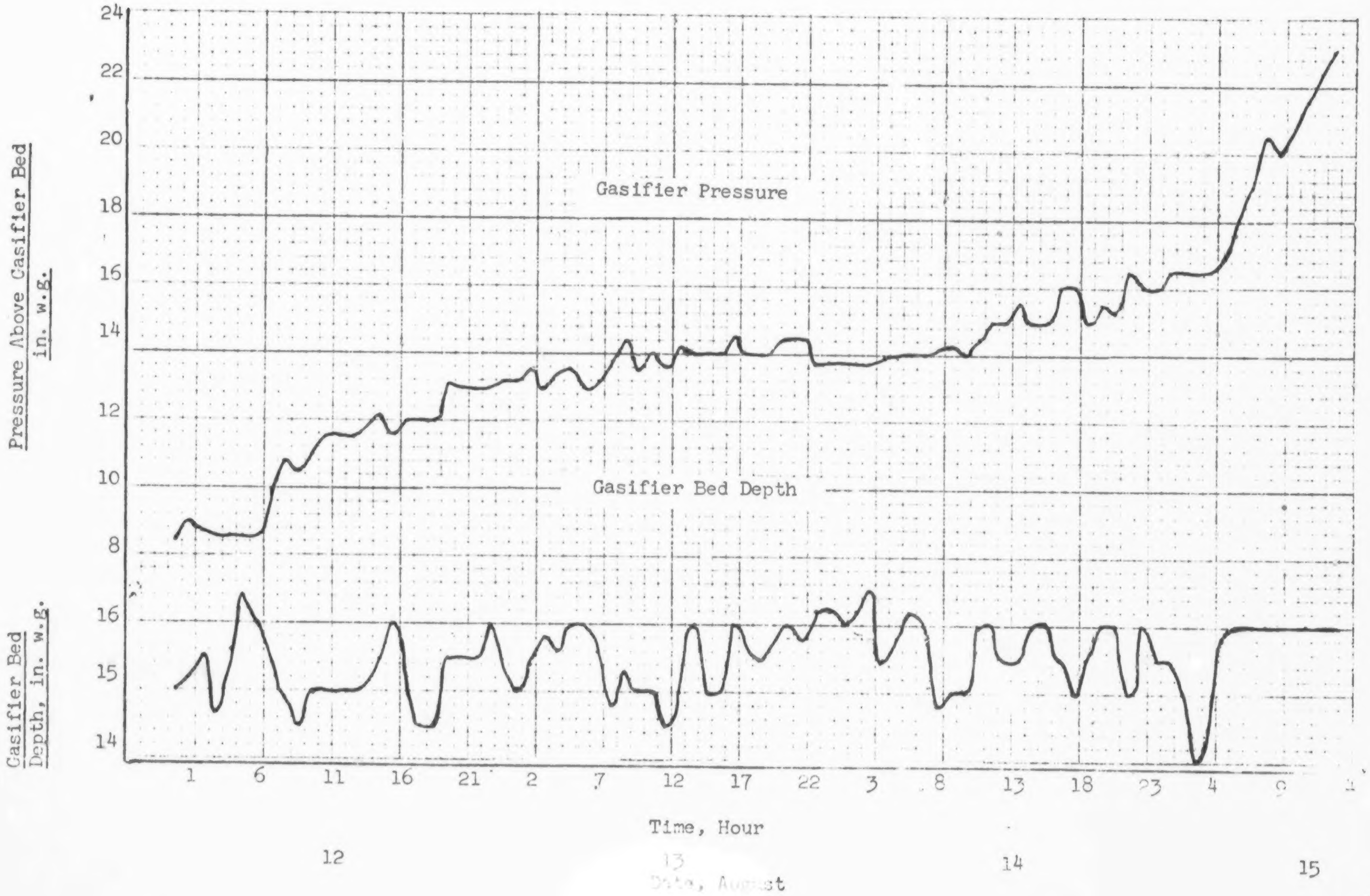


Fig I-3

CAFB Pilot Plant Operation Log (Run 1)

Regenerator and Gasifier Bed Temperatures

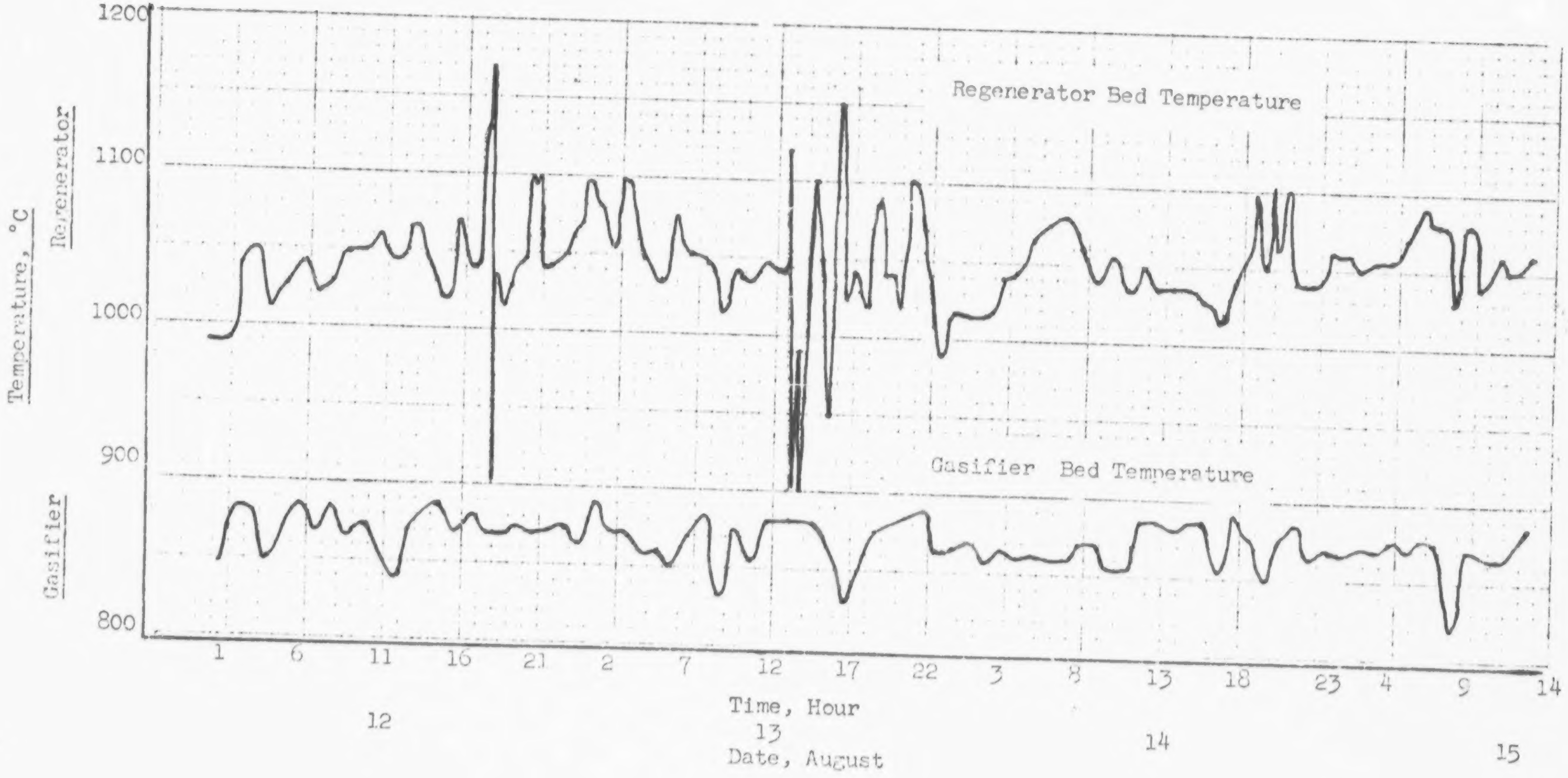


Fig J-4

CAFB Pilot Plant Operation Log (Run 1)

Regenerator Gas SO_2 Concentration and Regenerator Air Rate

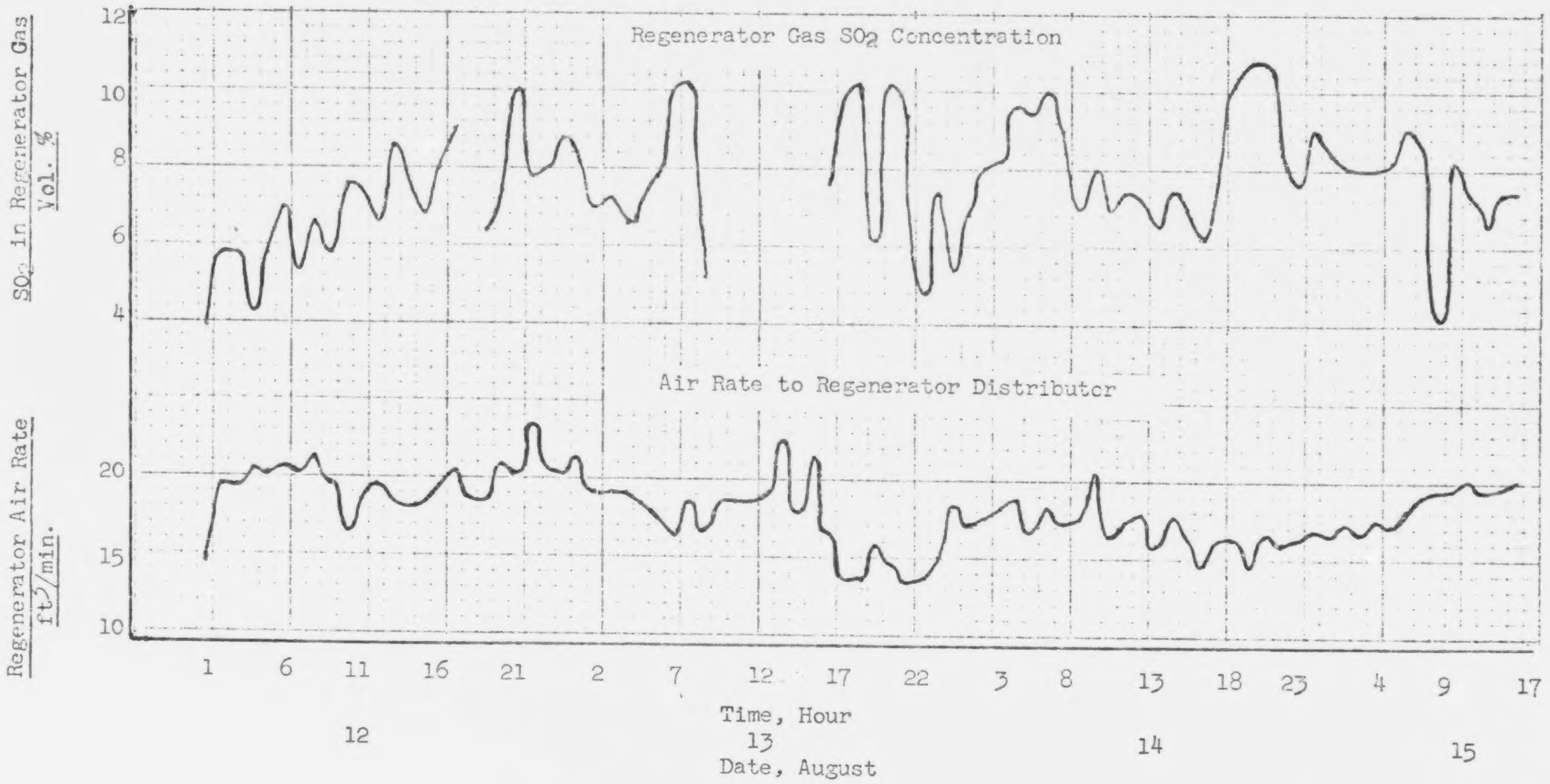


Fig J-5

CAFB Pilot Plant Operation Log (Run 1)

Regenerator Gas Oxygen and CO_2 Concentrations

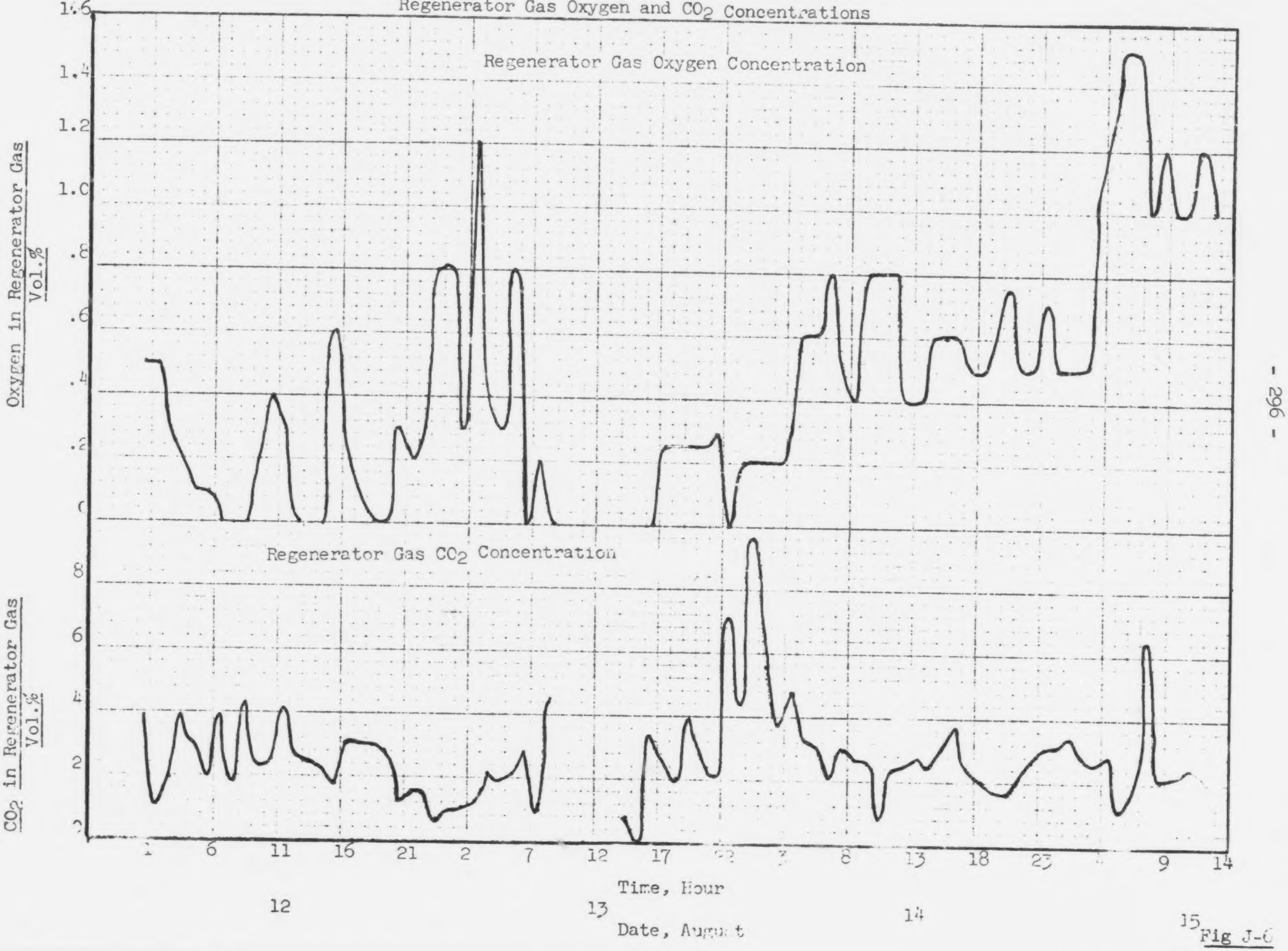


Fig J-6

CAFB Pilot Plant Operation Log (Run 1)
Selectivity of Oxidation of CaS to CaO

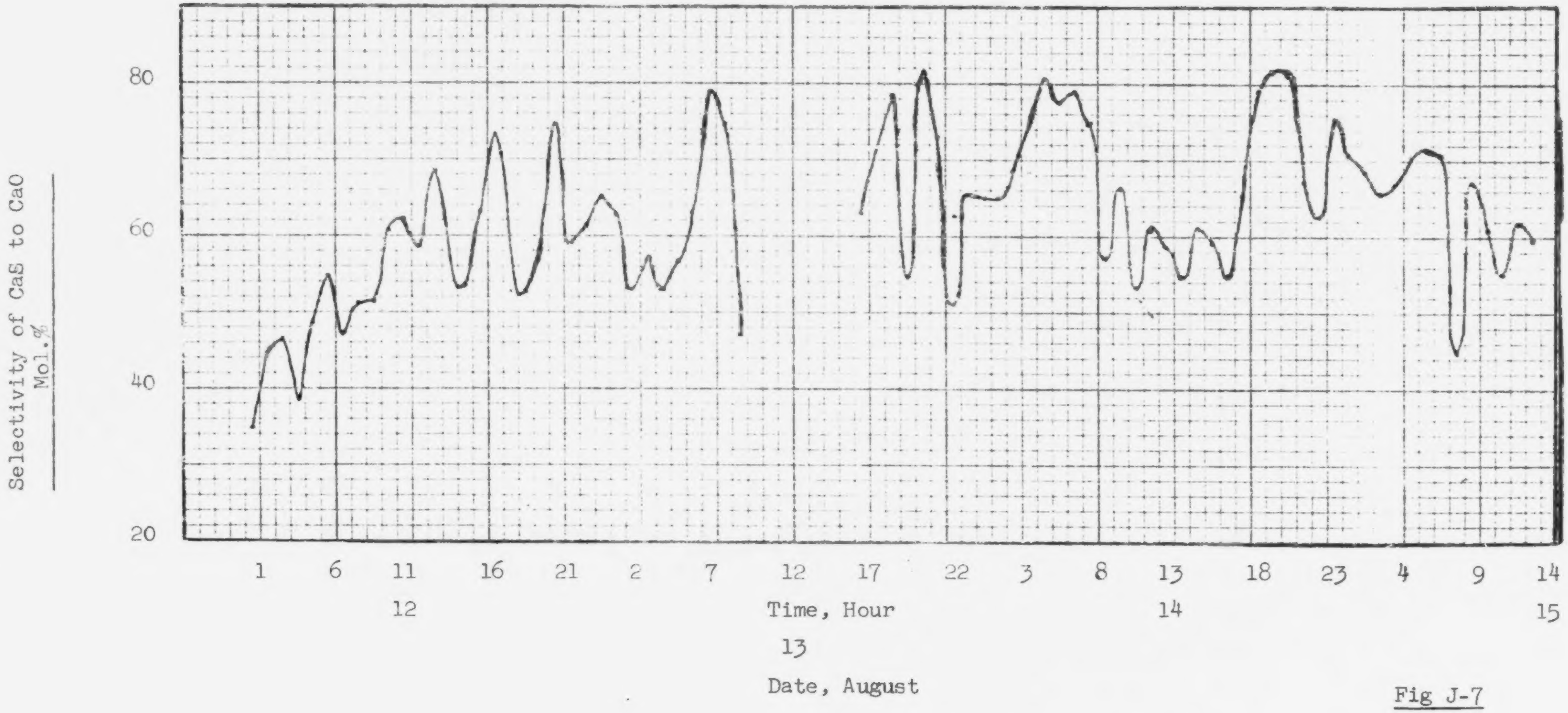
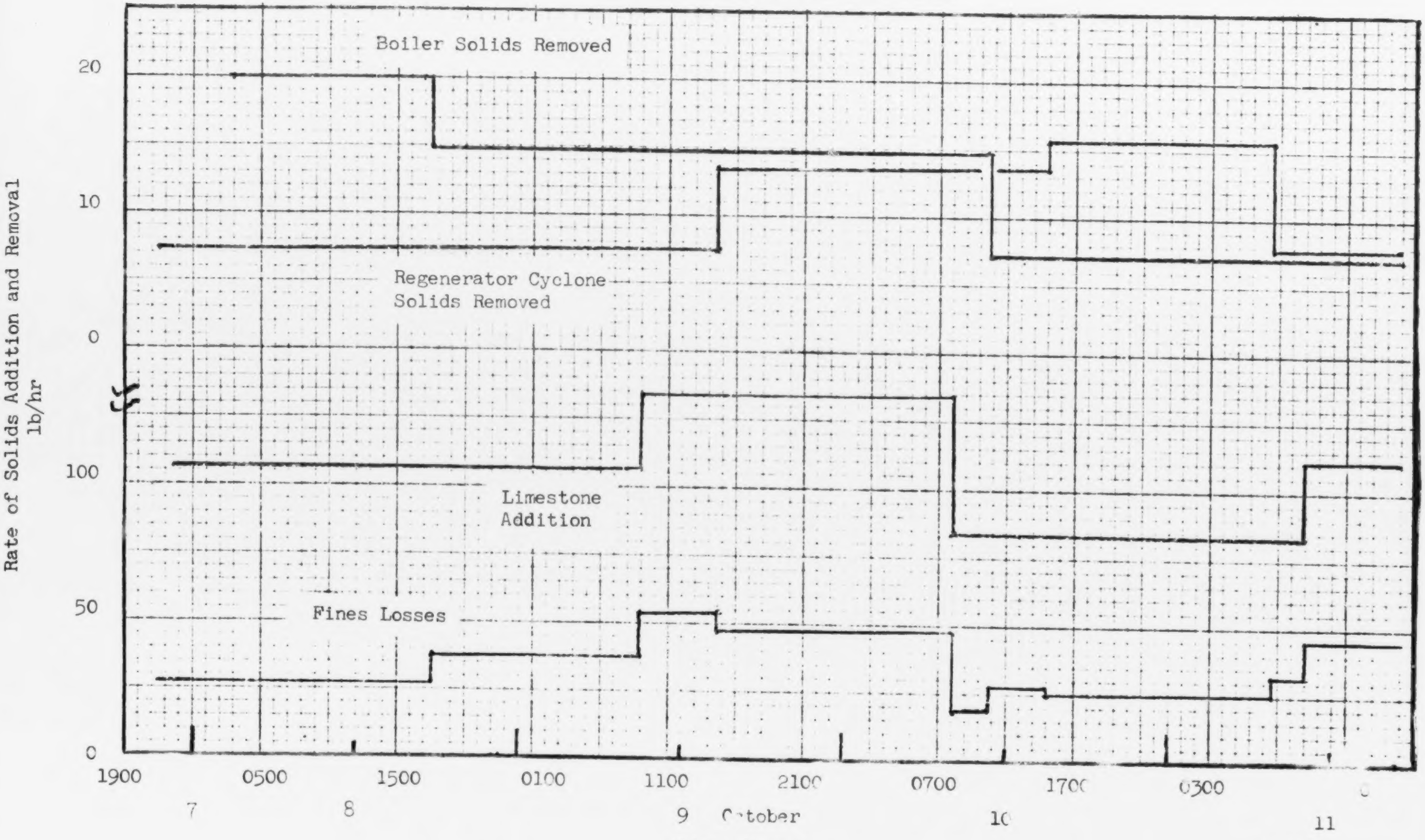


Fig J-7

CAFB Pilot Plant Operation Log (Run 2)

Solids Addition and Removal Rates



J-8

CAFB Pilot Plant Operation Log (Run 2)

Oil Feed and Calcium/Sulphur Feed Ratio

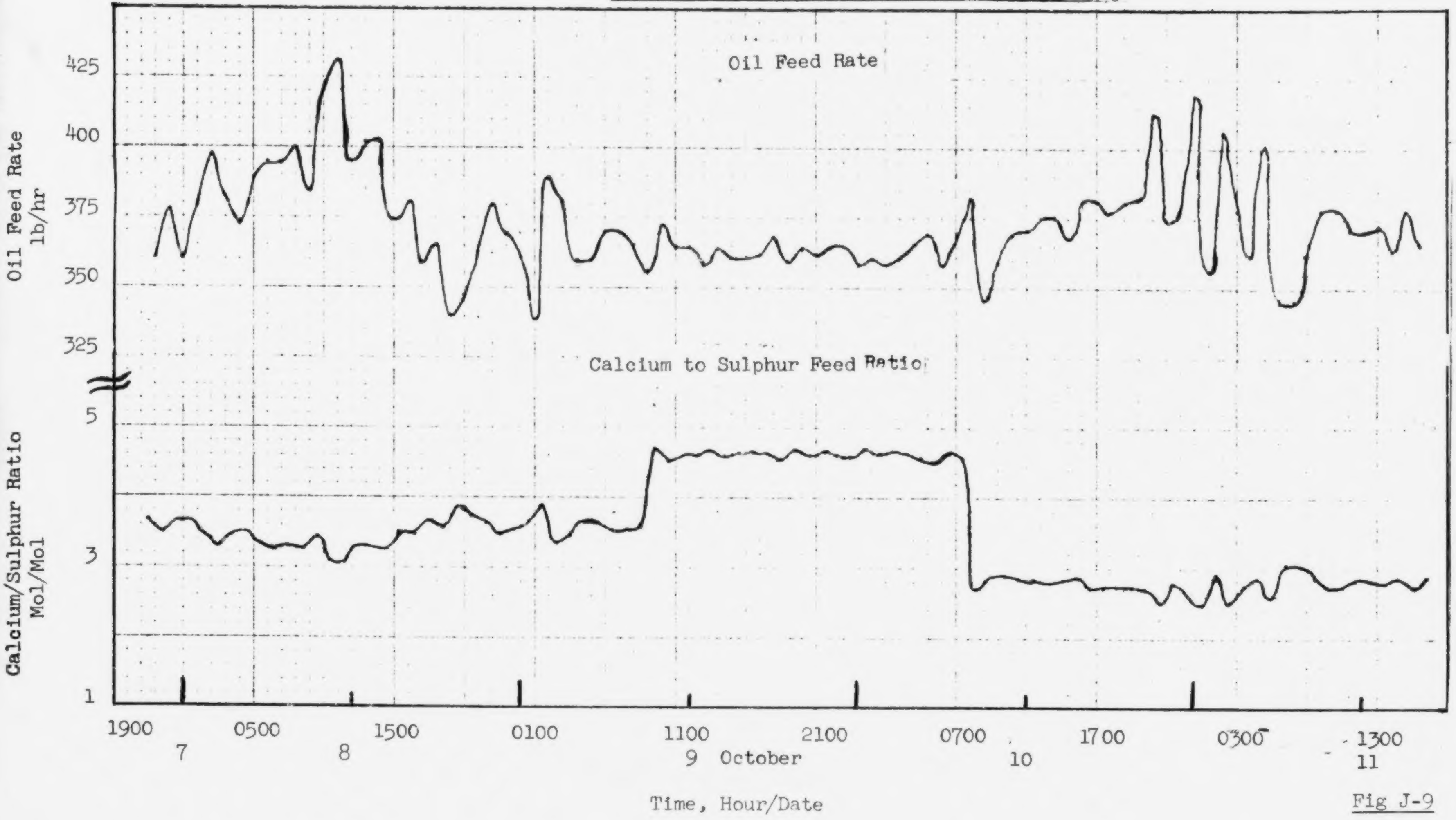


Fig J-9

CAFB Pilot Plant Operation Log (Run 2)

Gasifier Air/Fuel Ratio and Gas Velocity

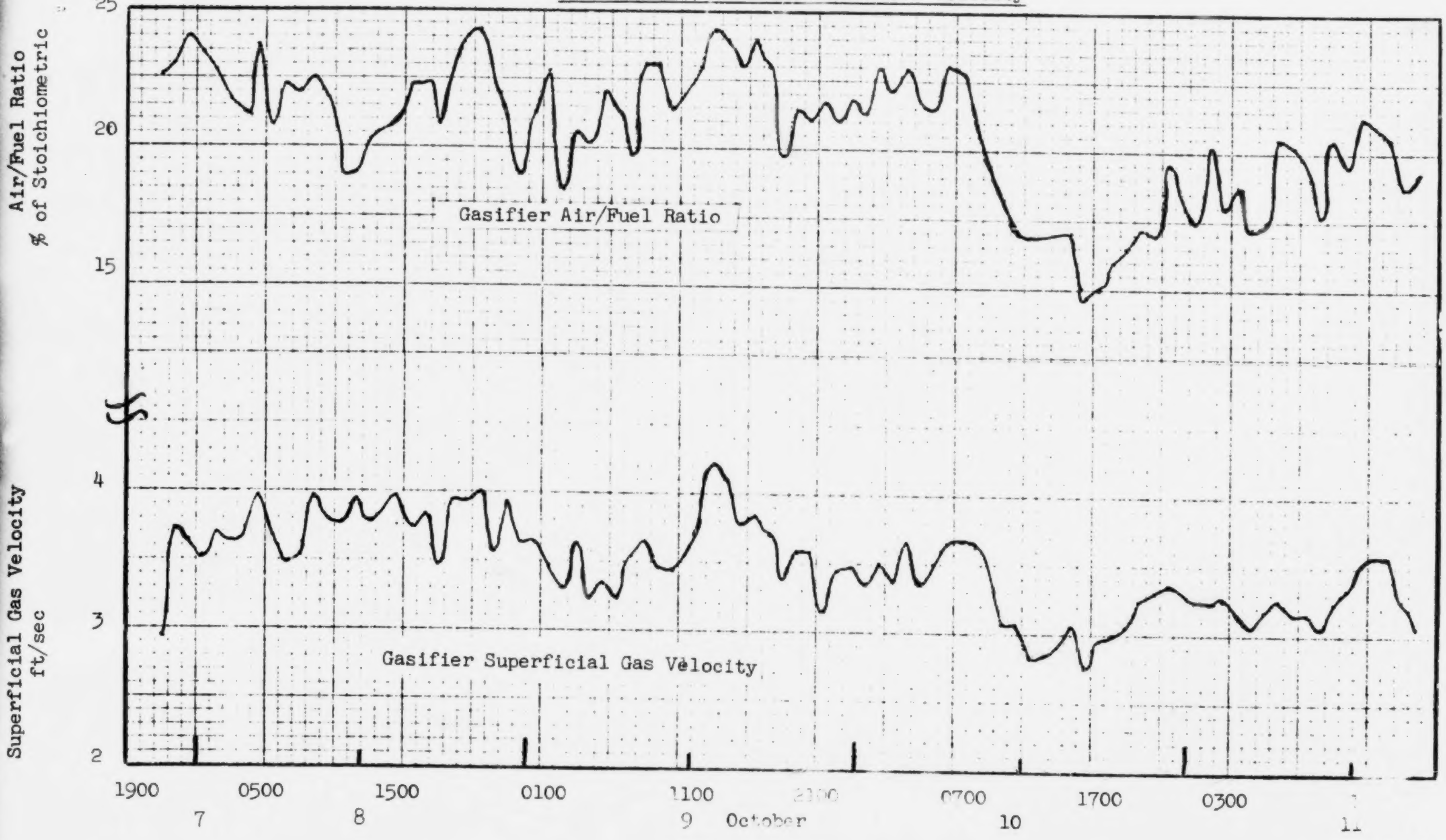


Fig J-10

CAFB Pilot Plant Operation Log (Run 2)

Gasifier Pressure and Bed Depth

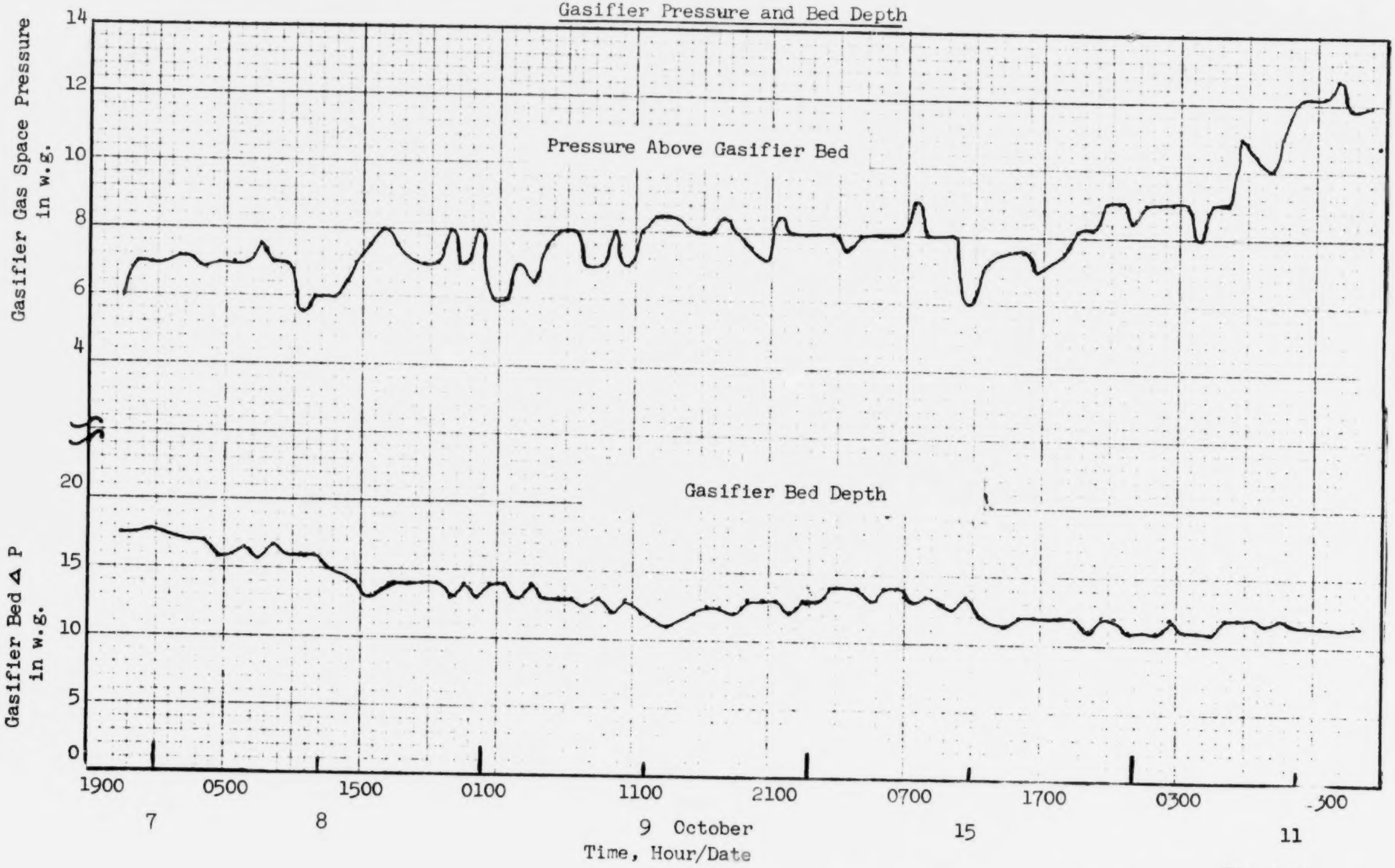


Fig J-11

CAFB Pilot Plant Operation Log (Run 2)

Regenerator and Gasifier Bed Temperatures

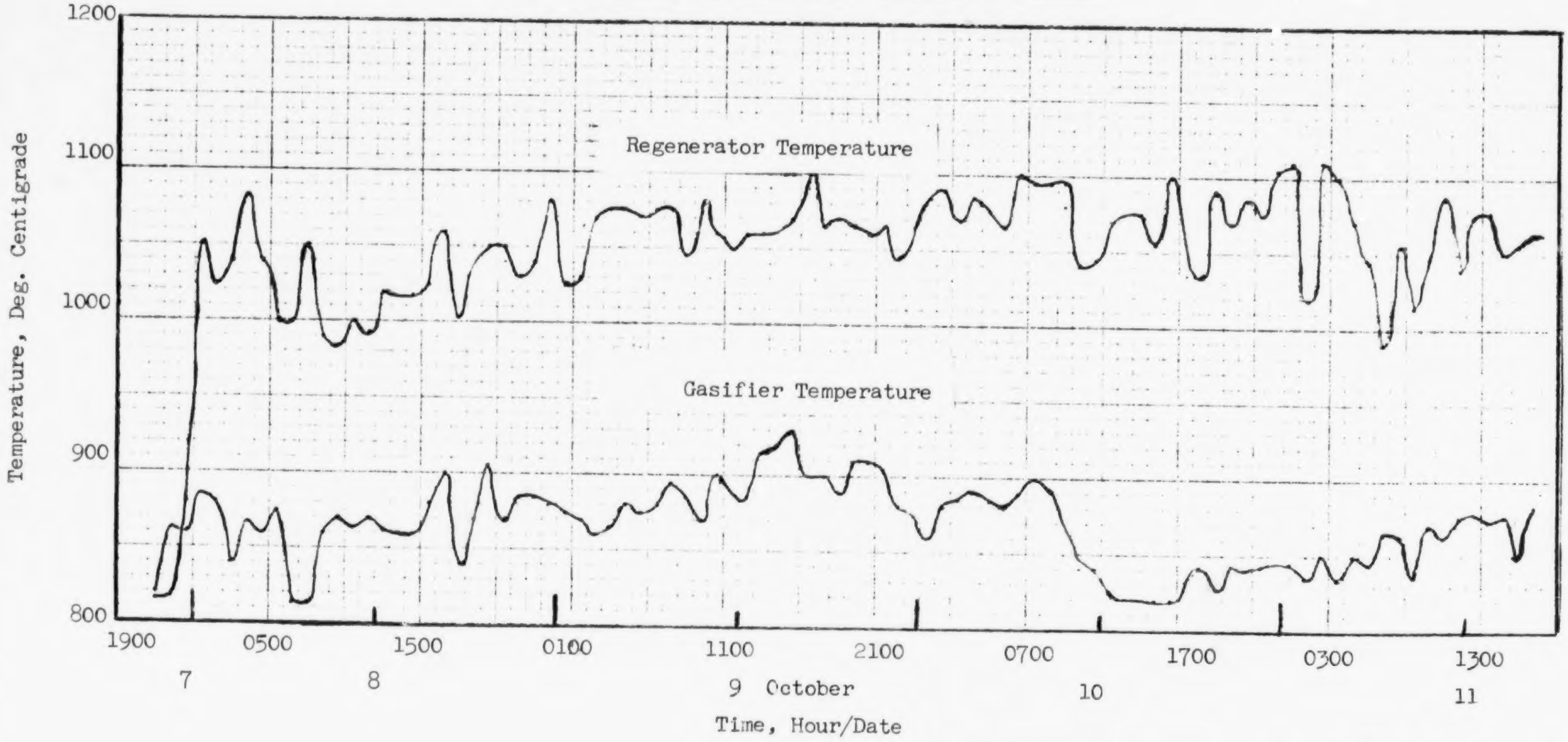


Fig J-12

CAFB Pilot Plant Operation Log (Run 2)

Regenerator Air Rate and SO_2 Concentration

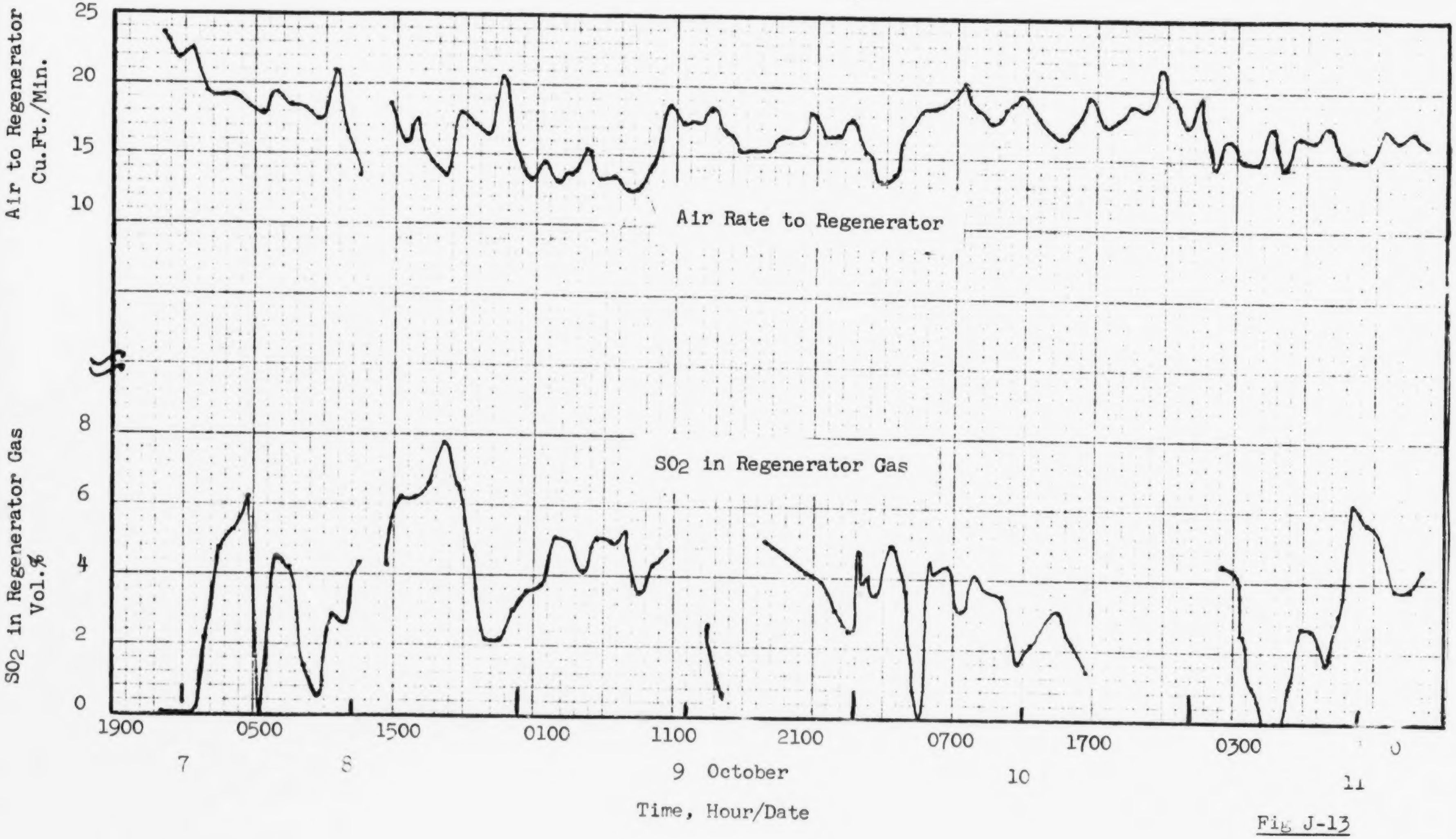


Fig J-13

CAFB Pilot Plant Operation Log (Run 2)

Regenerator Gas Oxygen and CO_2 Concentration

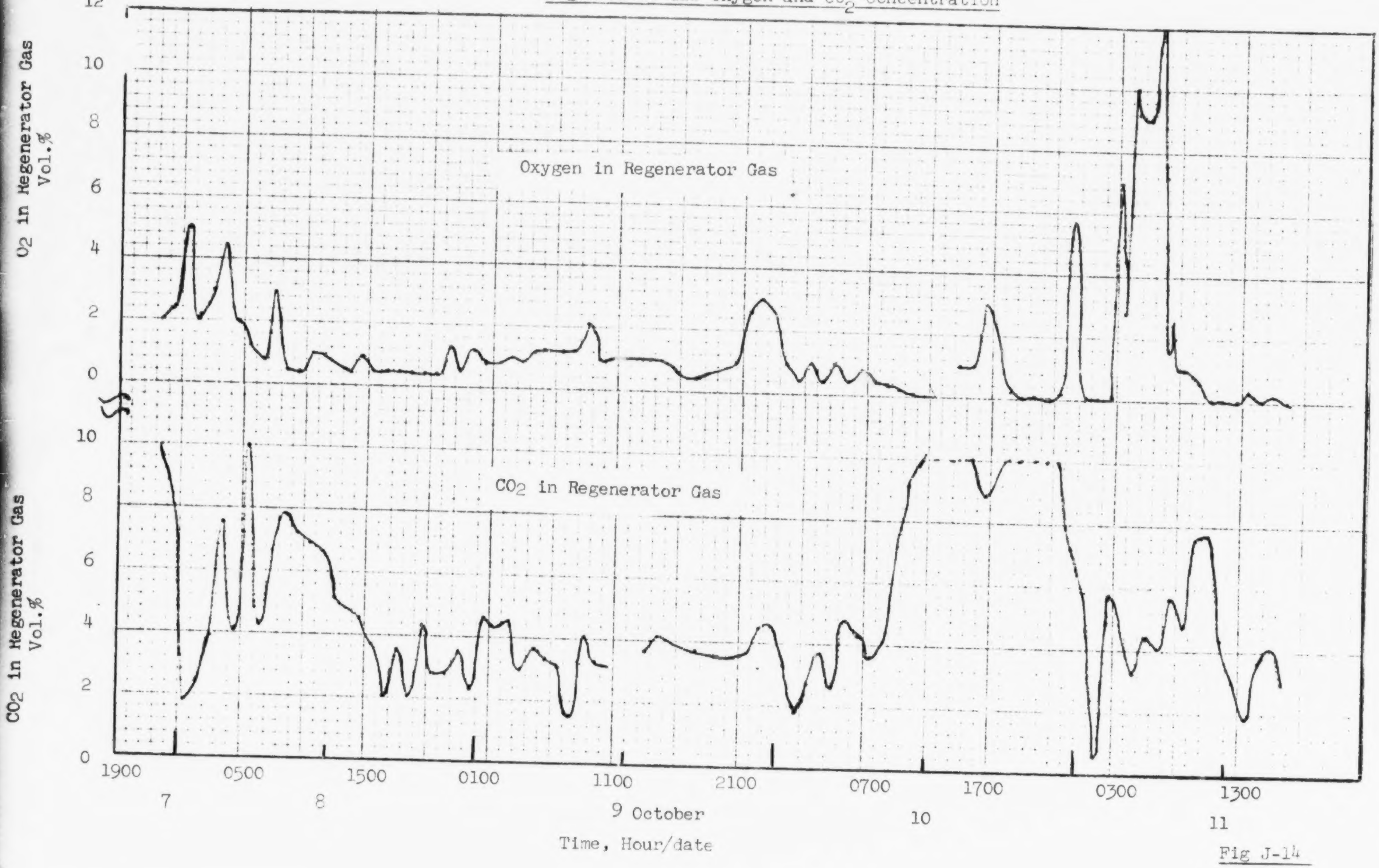


Fig J-14

CAFB Pilot Plant Operation Log (Run 2)

Selectivity of CaS Oxidation and SO_2 Output of Regenerator

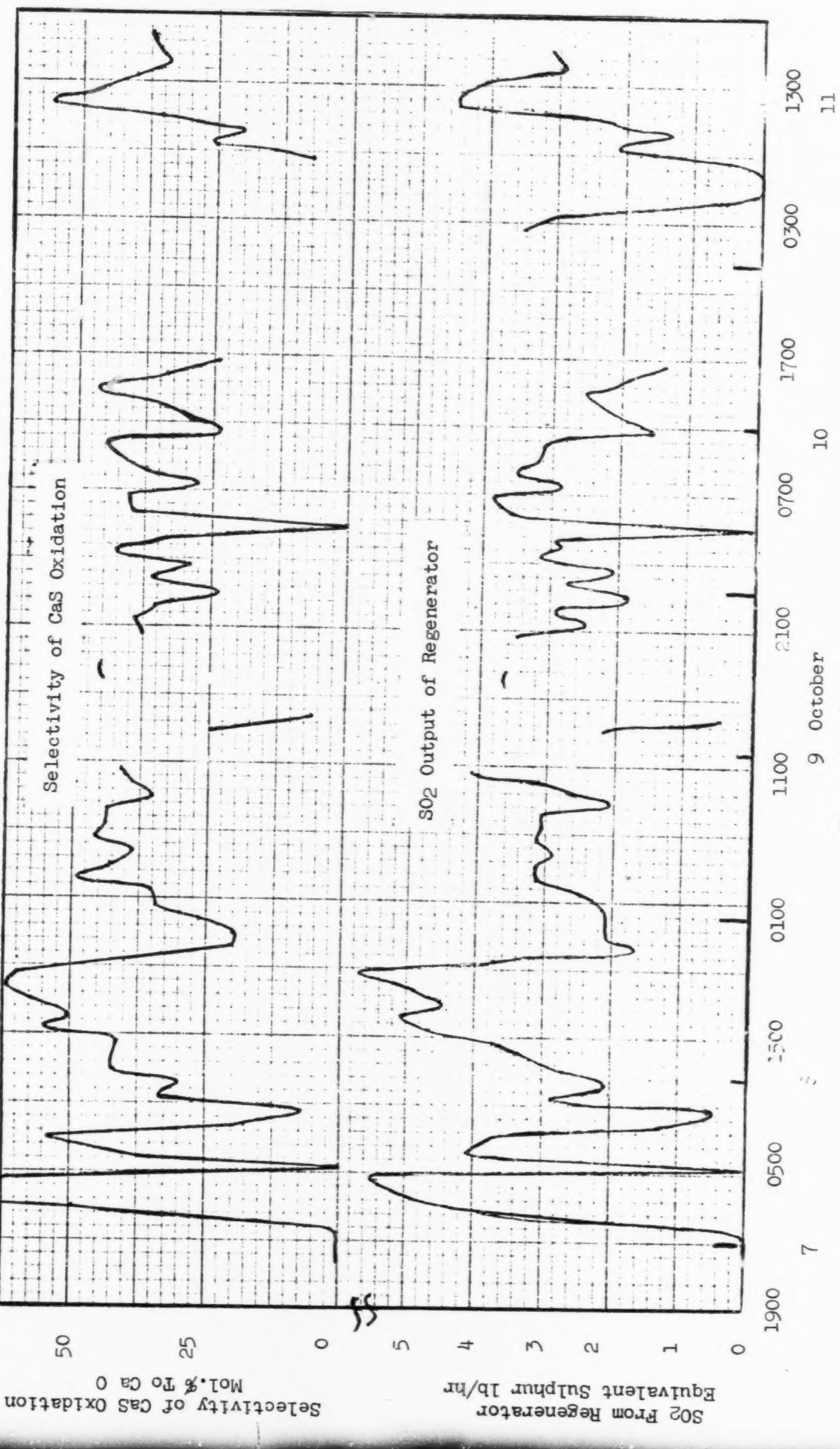


Fig.J-15

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

SULPHUR REMOVAL

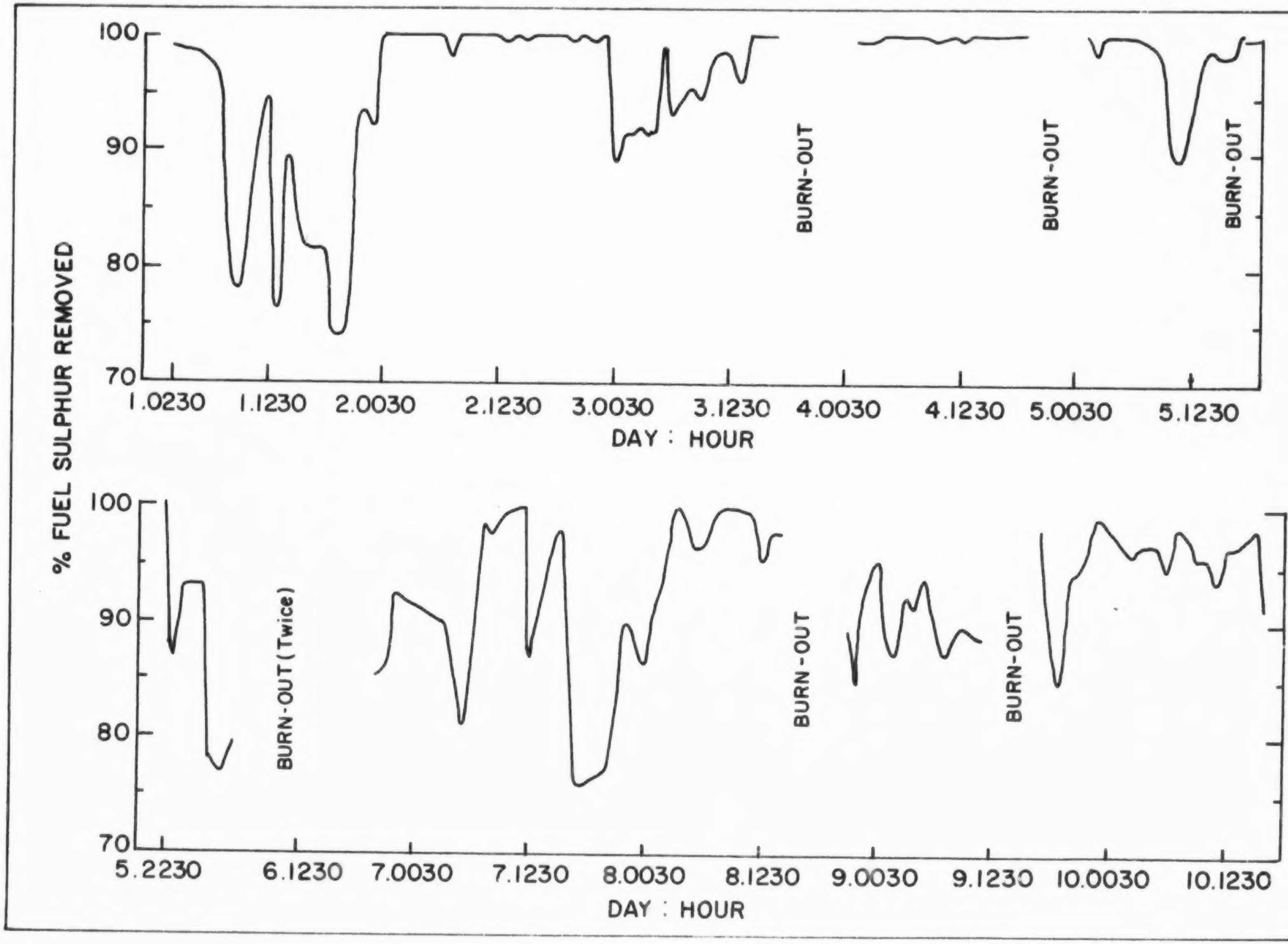


FIG. J-16

CONTINUOUS PILOT PLANT OPERATION LOG(RUN.3.)

OIL RATE

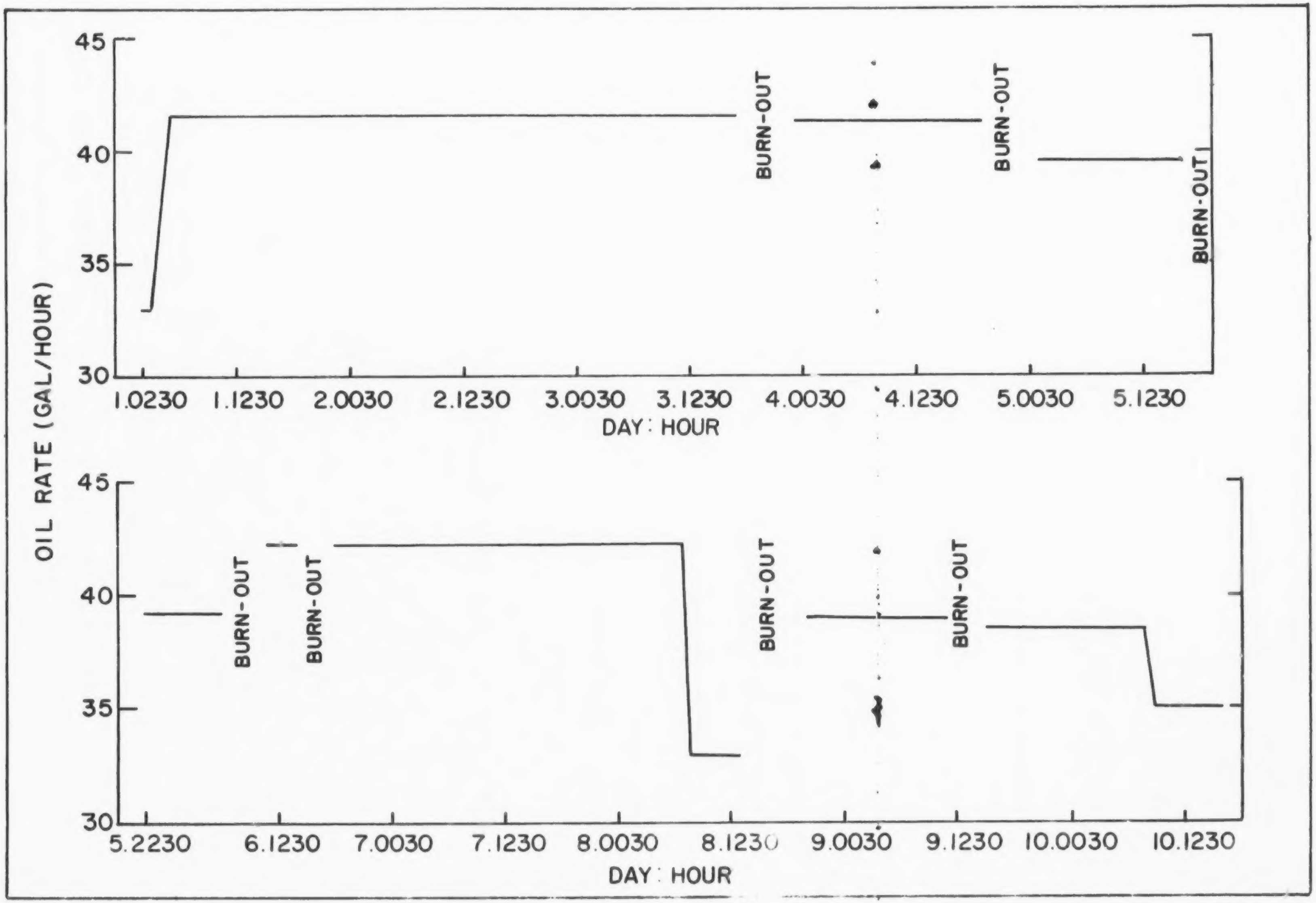


FIG. J-17.

CAFB PILOT PLANT OPERATIONS LOG (RUN. 3.)

CUMULATIVE LIME ADDITION AND REGENERATION SOLIDS REMOVAL

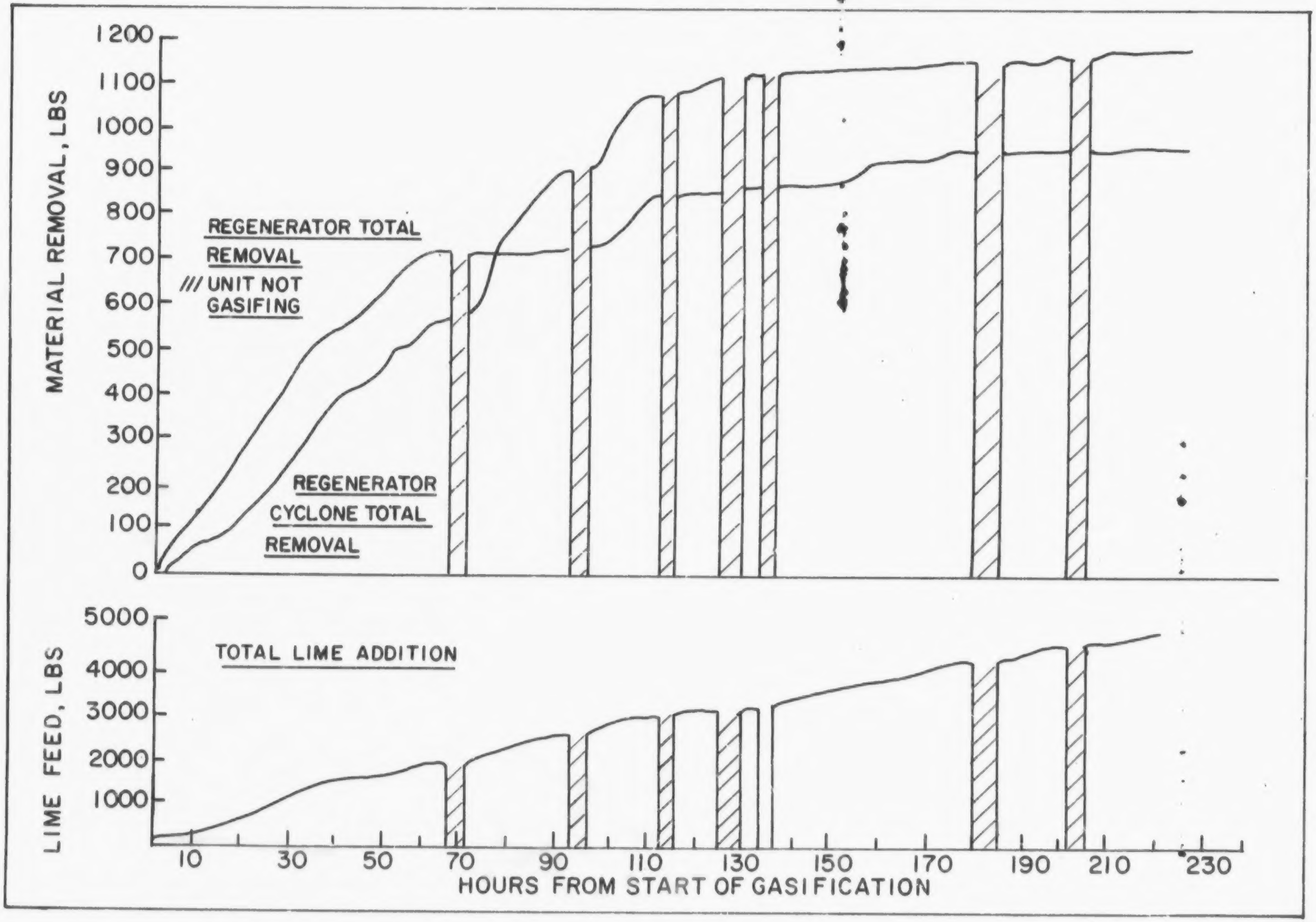


FIG. J-18

CAFB PILOT PLANT OPERATIONS LOG (RUN.3.)

CUMULATIVE SOLIDS REMOVAL FROM BOILER AND FLUE

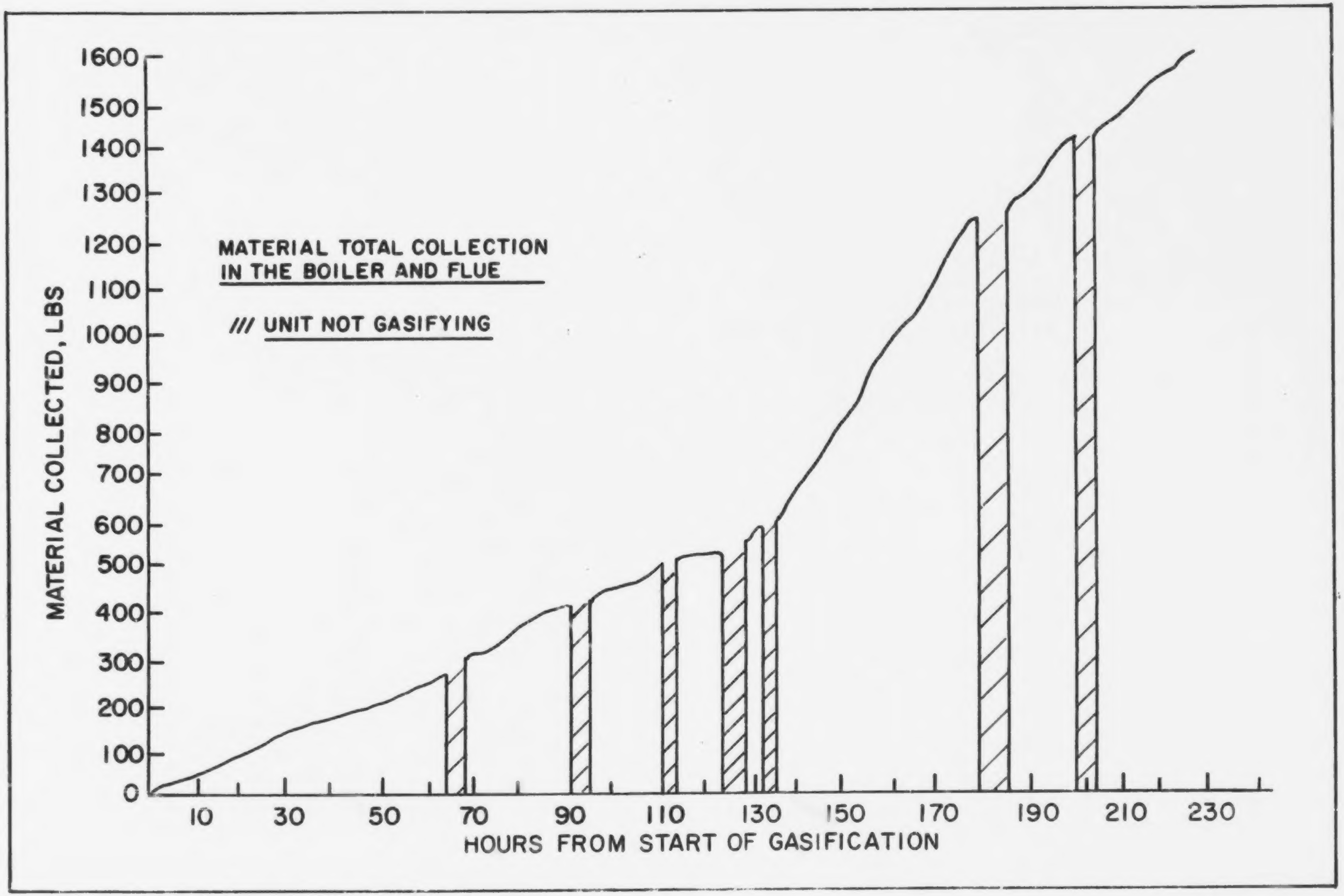


FIG. J-19

CAFB PILOT PLANT OPERATIONS LOG (RUN.3.)

RATES OF LIME FEED AND REGENERATOR SOLIDS REMOVAL

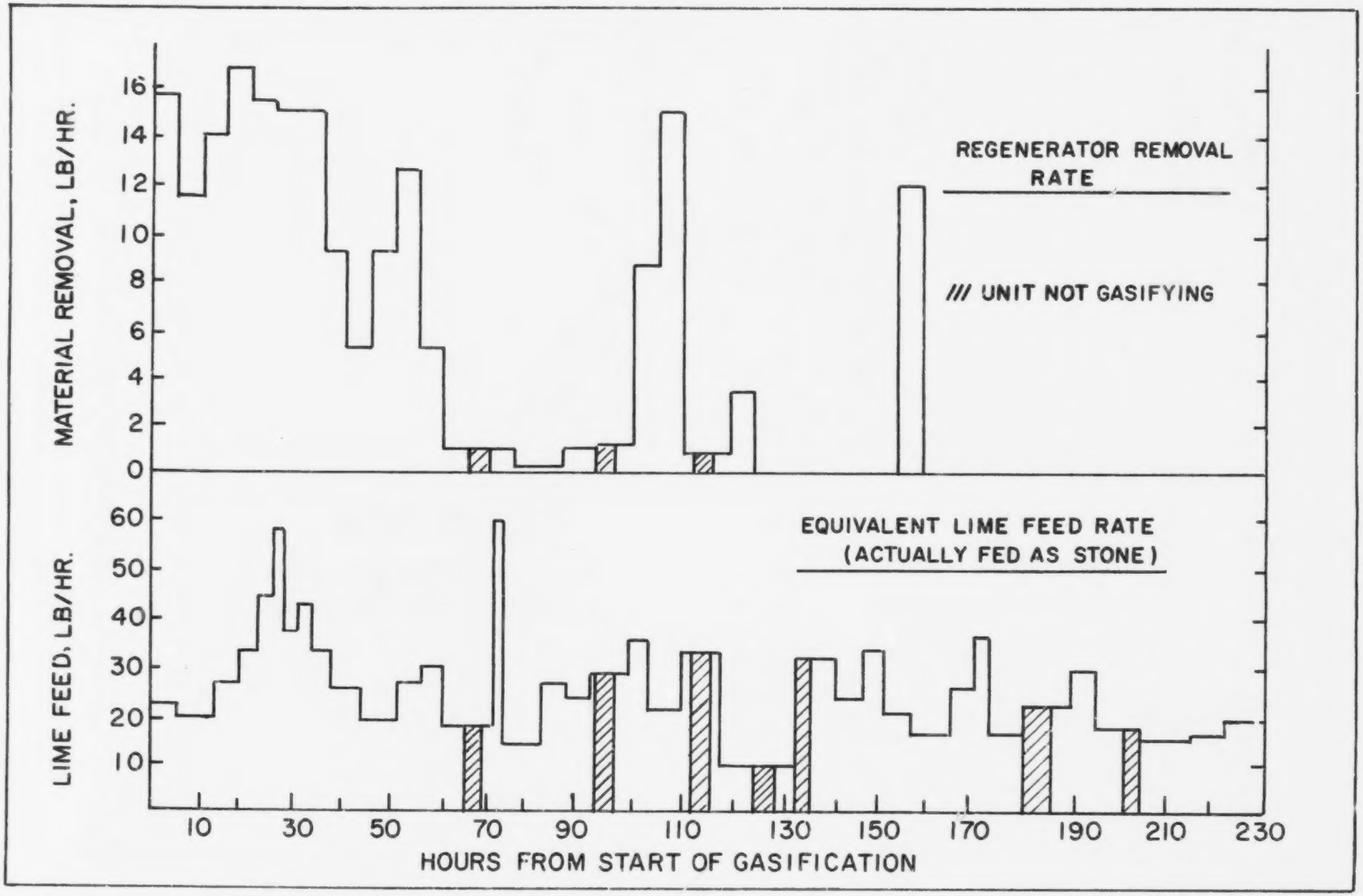


FIG. J-20

CAFB PILOT PLANT OPERATIONS LOG (RUN.3.)
REGENERATOR CYCLONE AND BOILER FLUE SOLIDS REMOVAL RATES

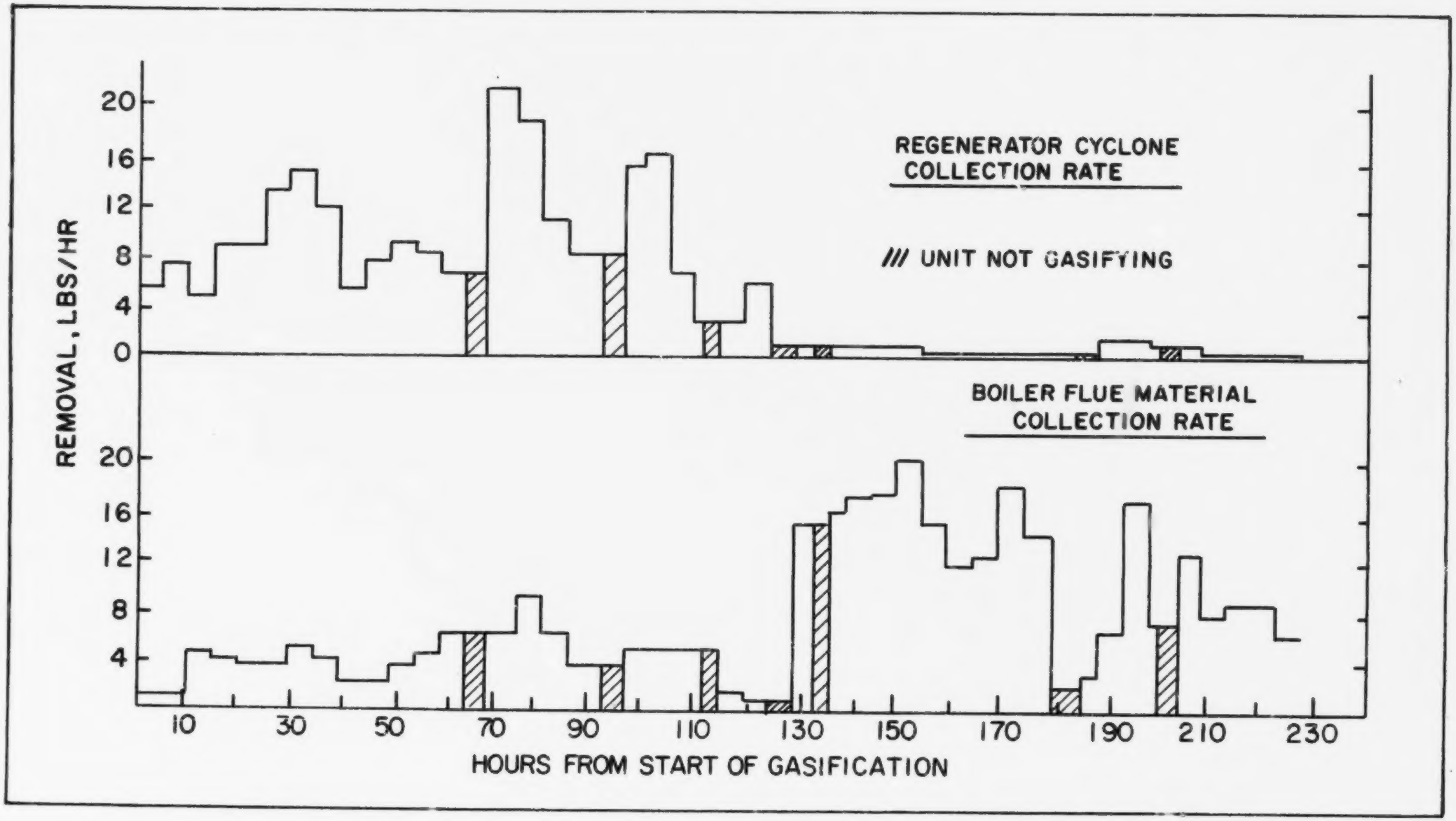


FIG. J-21

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

BED REPLACEMENT RATE

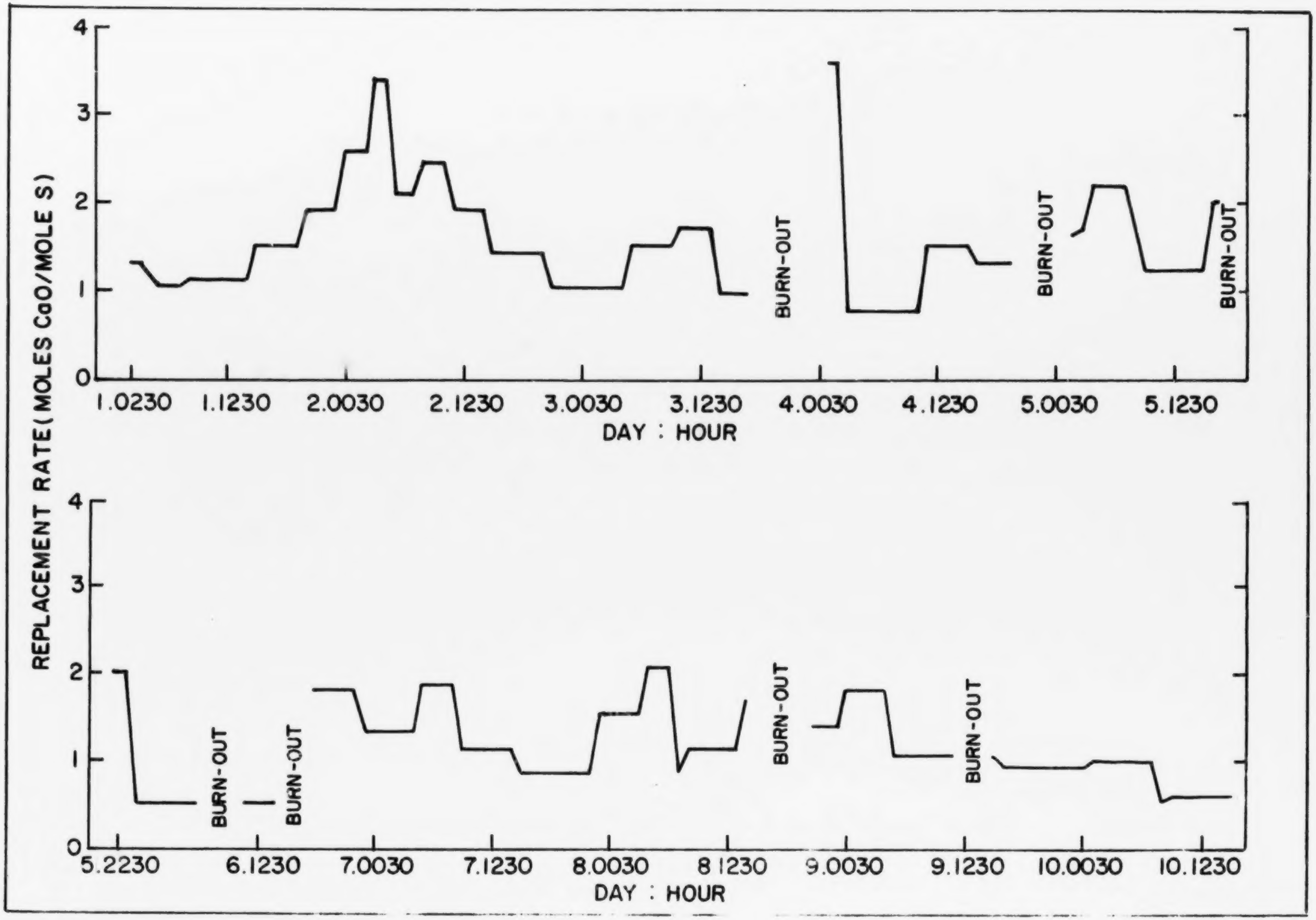


FIG. J-22.

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

AIR/FUEL RATIO

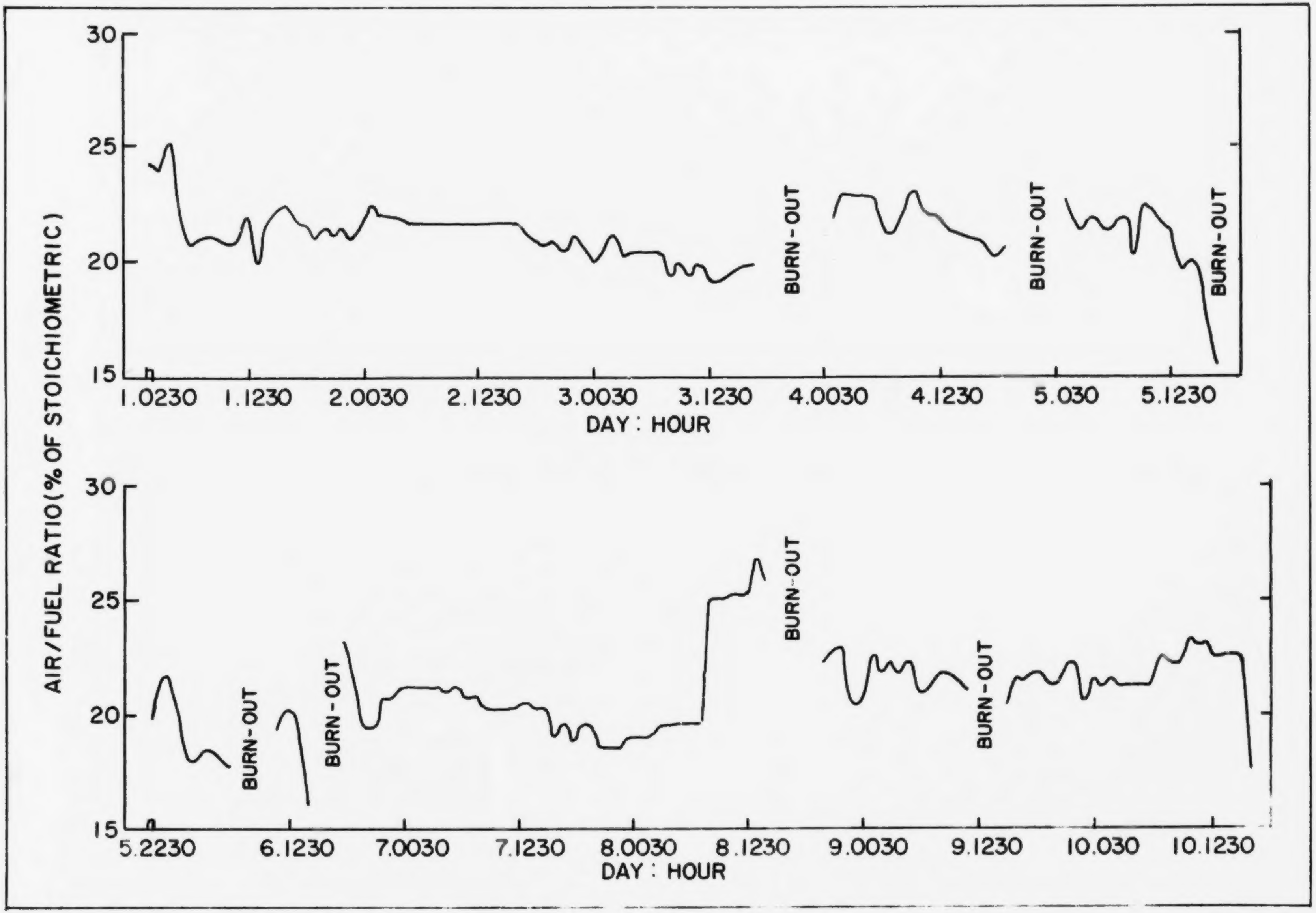


FIG. J-23.

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

SUPERFICIAL GAS VELOCITY

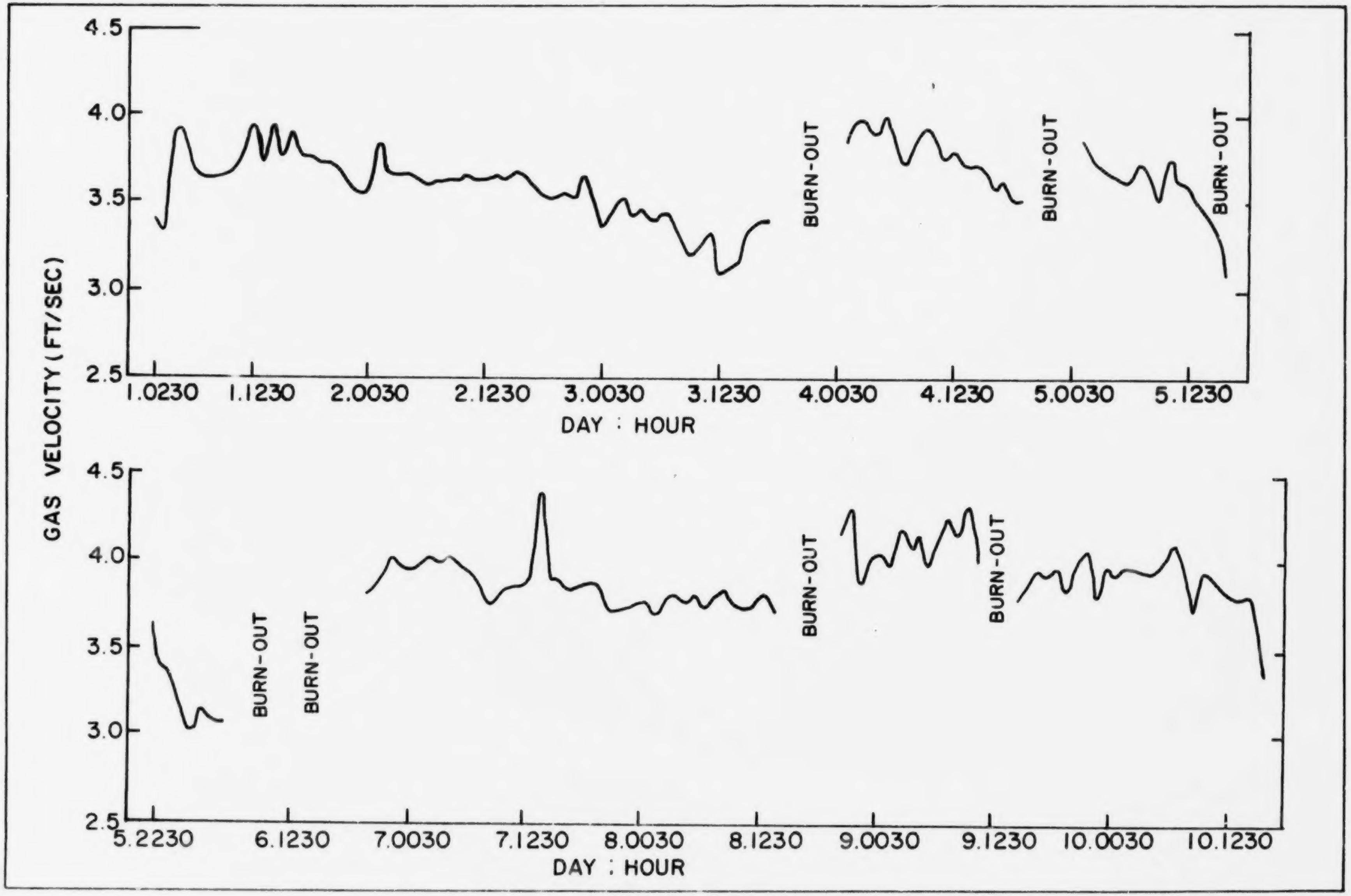


FIG. J-24

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

GASIFIER GAS SPACE PRESSURE

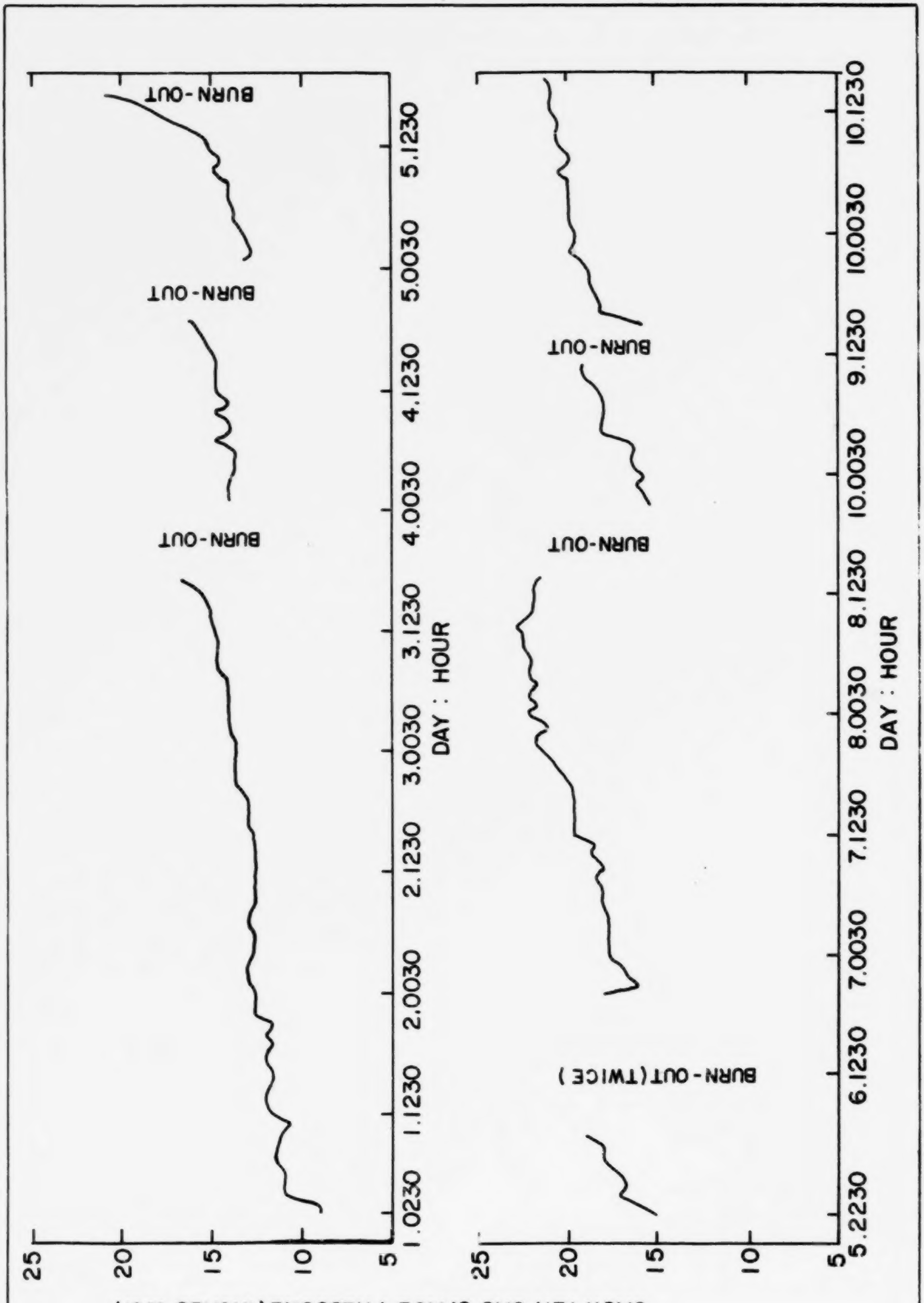


FIG. J-25

CAFB PILOT PLANT OPERATION LOG (RUN. 3.)

DISTRIBUTOR PRESSURE DROP

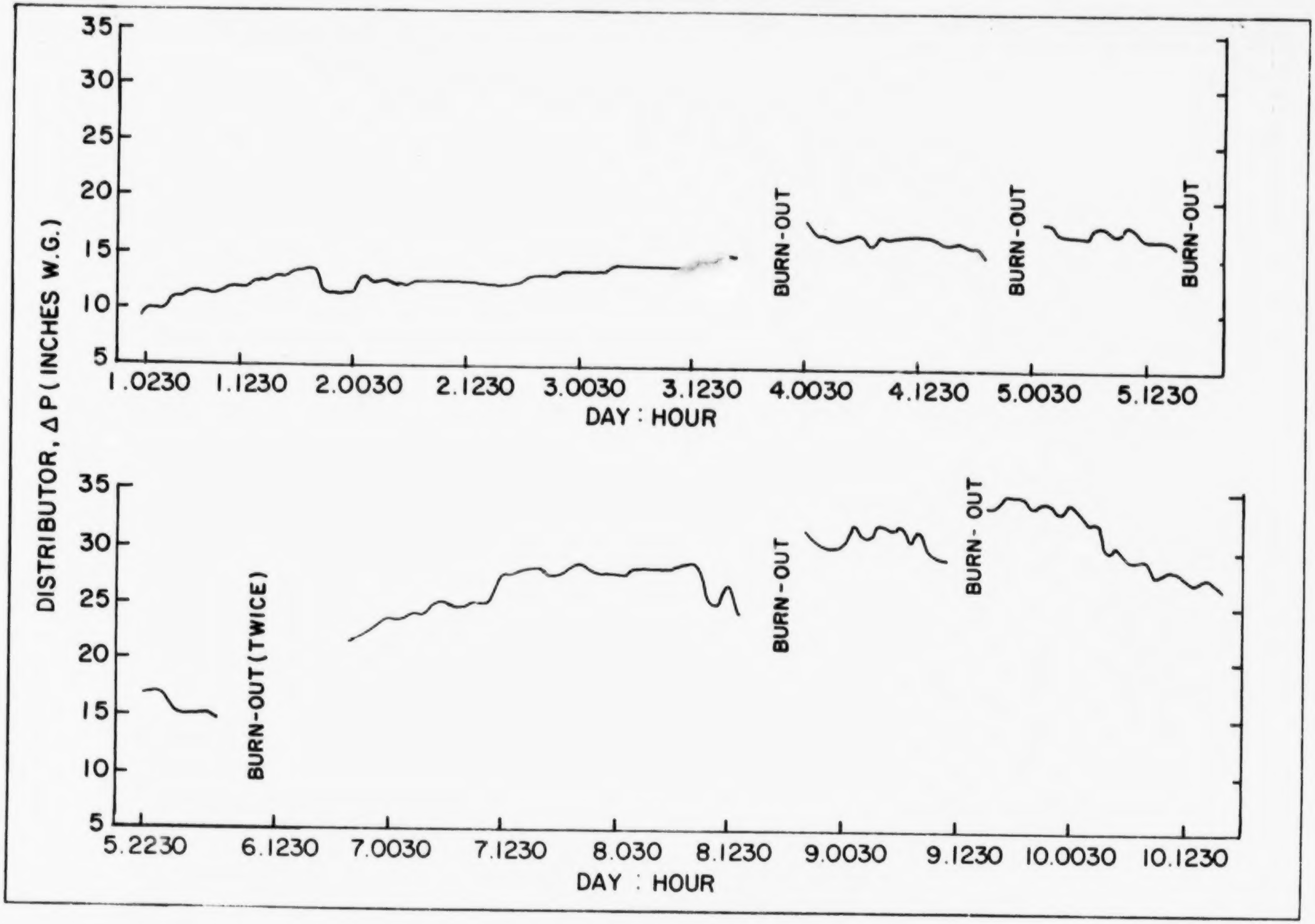
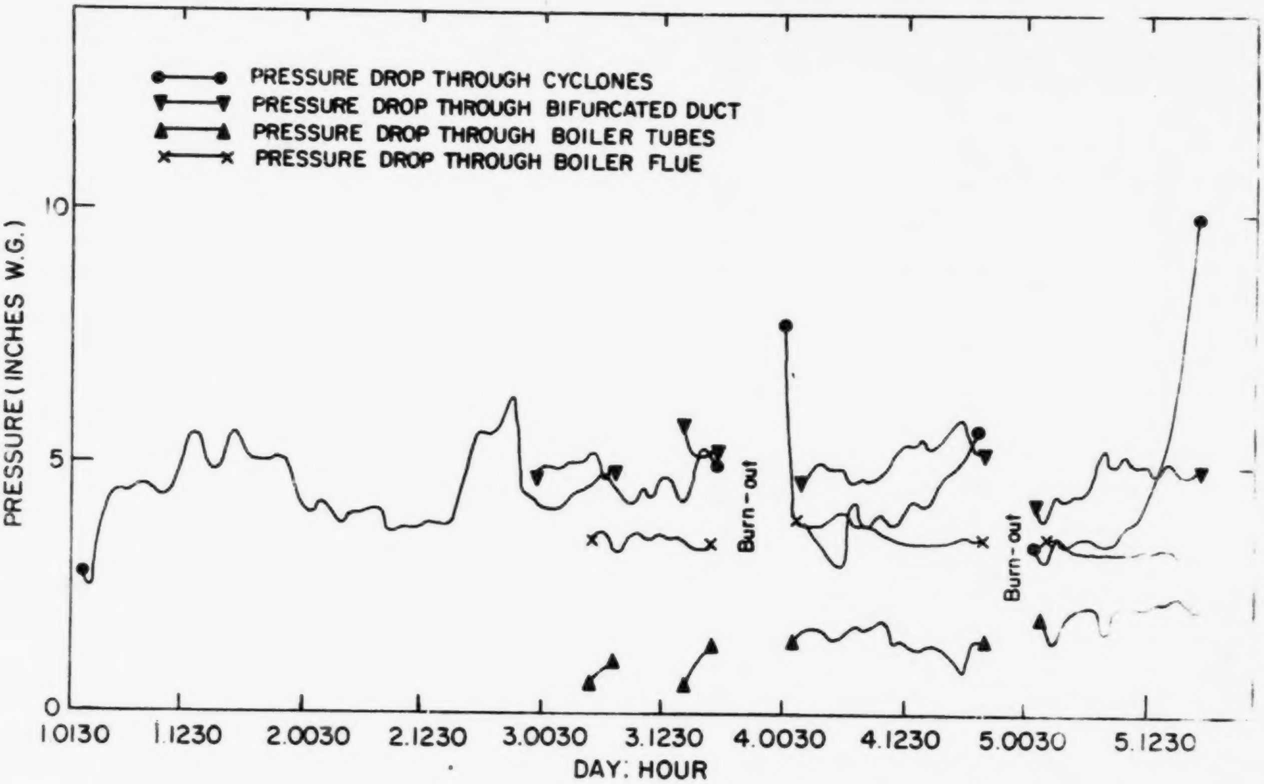


FIG. J-26

CAFB PILOT PLANT OPERATIONS LOG(RUN.3)

PRESSURE DROPS DOWNSTREAM OF GASIFIER



PRESSURE DROPS DOWNSTREAM OF GASIFIER

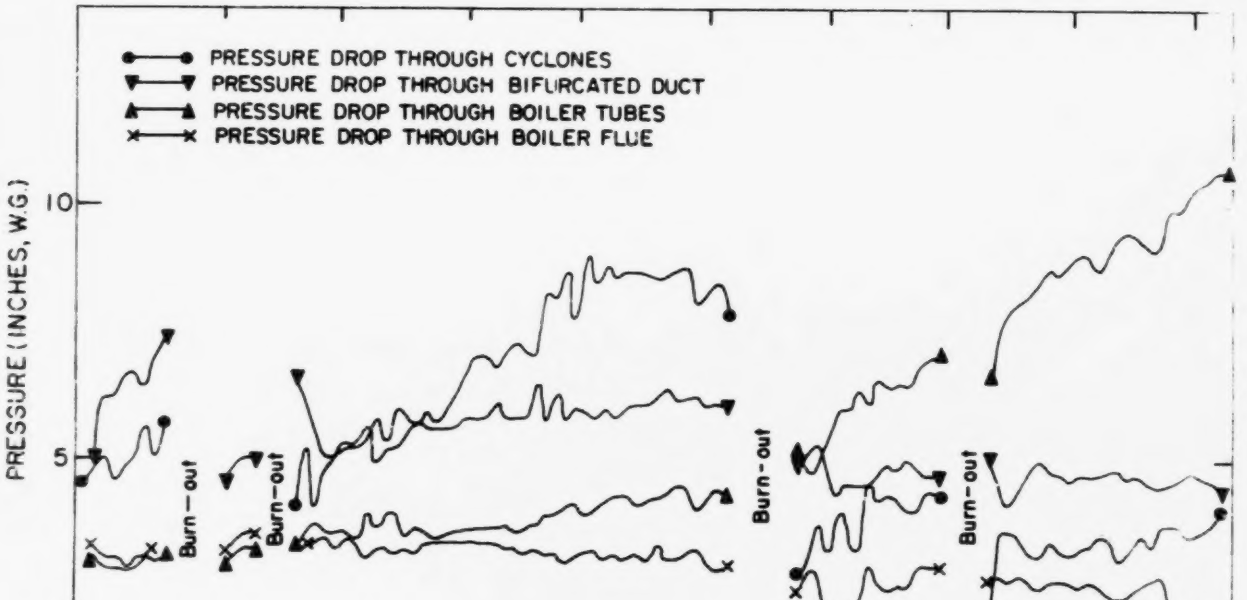


FIG. J27

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

GASIFIER BED DEPTH

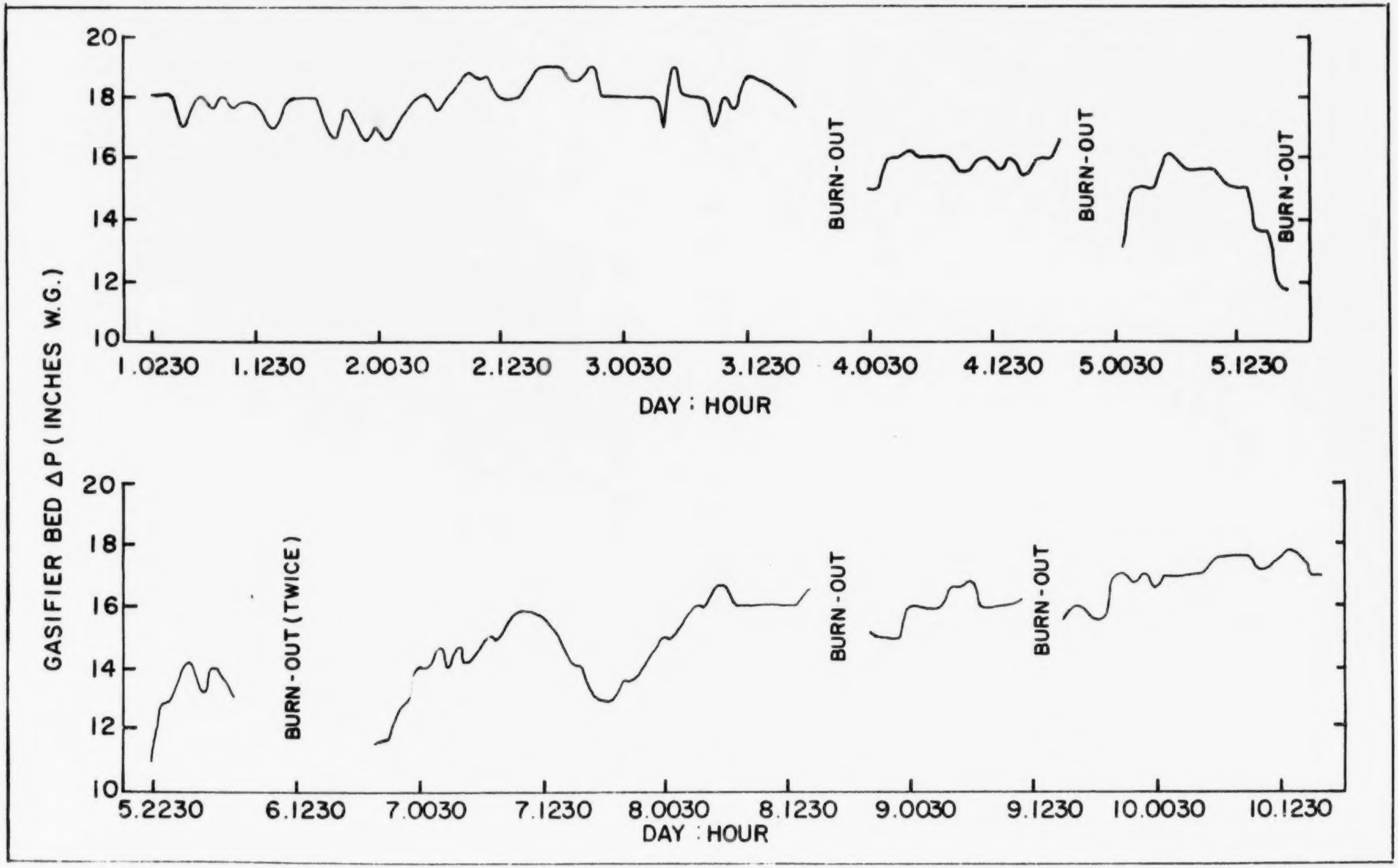


FIG. J-28.

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

BED TEMPERATURES

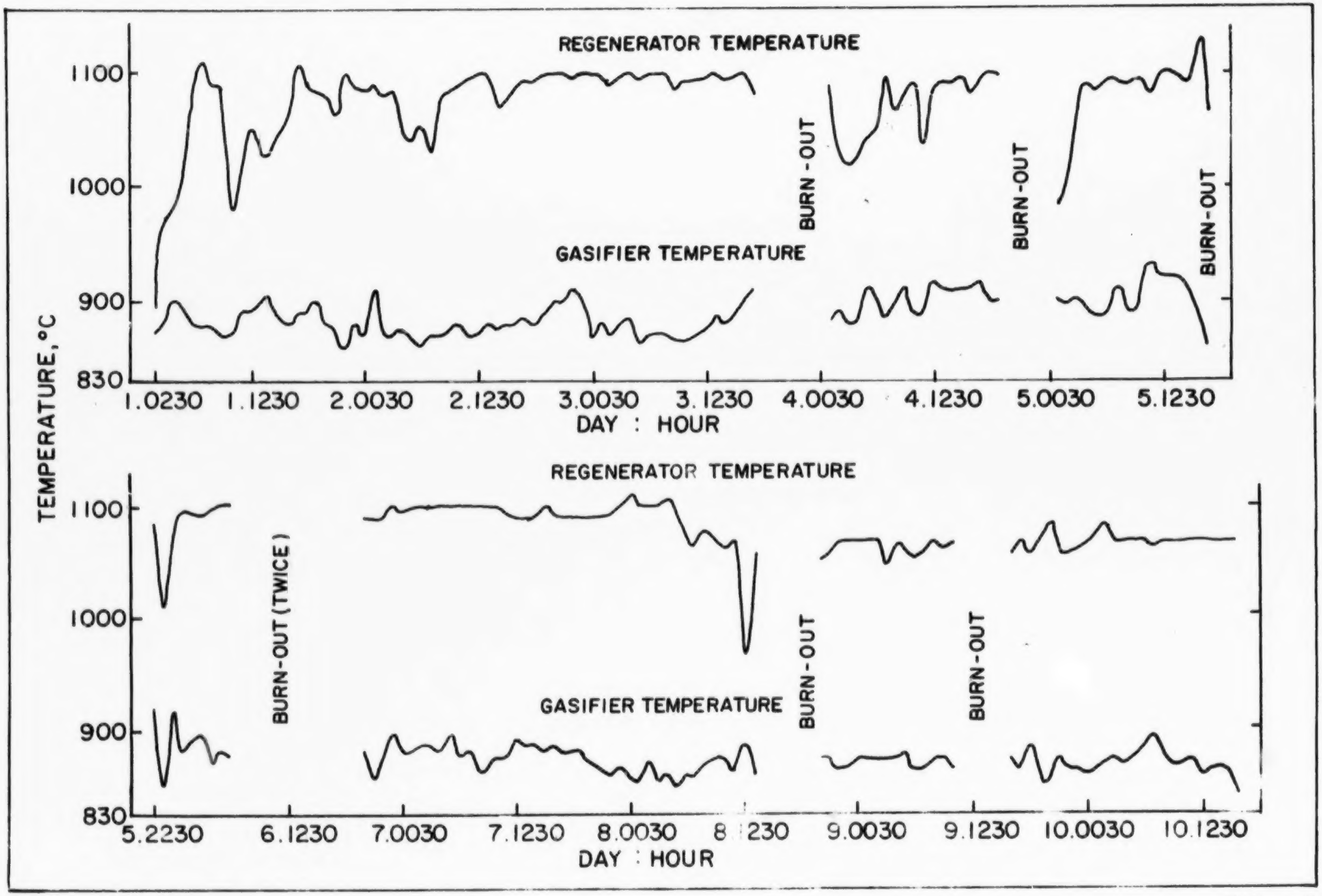


FIG J-29

CONTINUOUS PILOT PLANT OPERATION LOG(RUN.3.)
GASIFIER TO REGENERATOR SOLIDS TRANSFER RATE

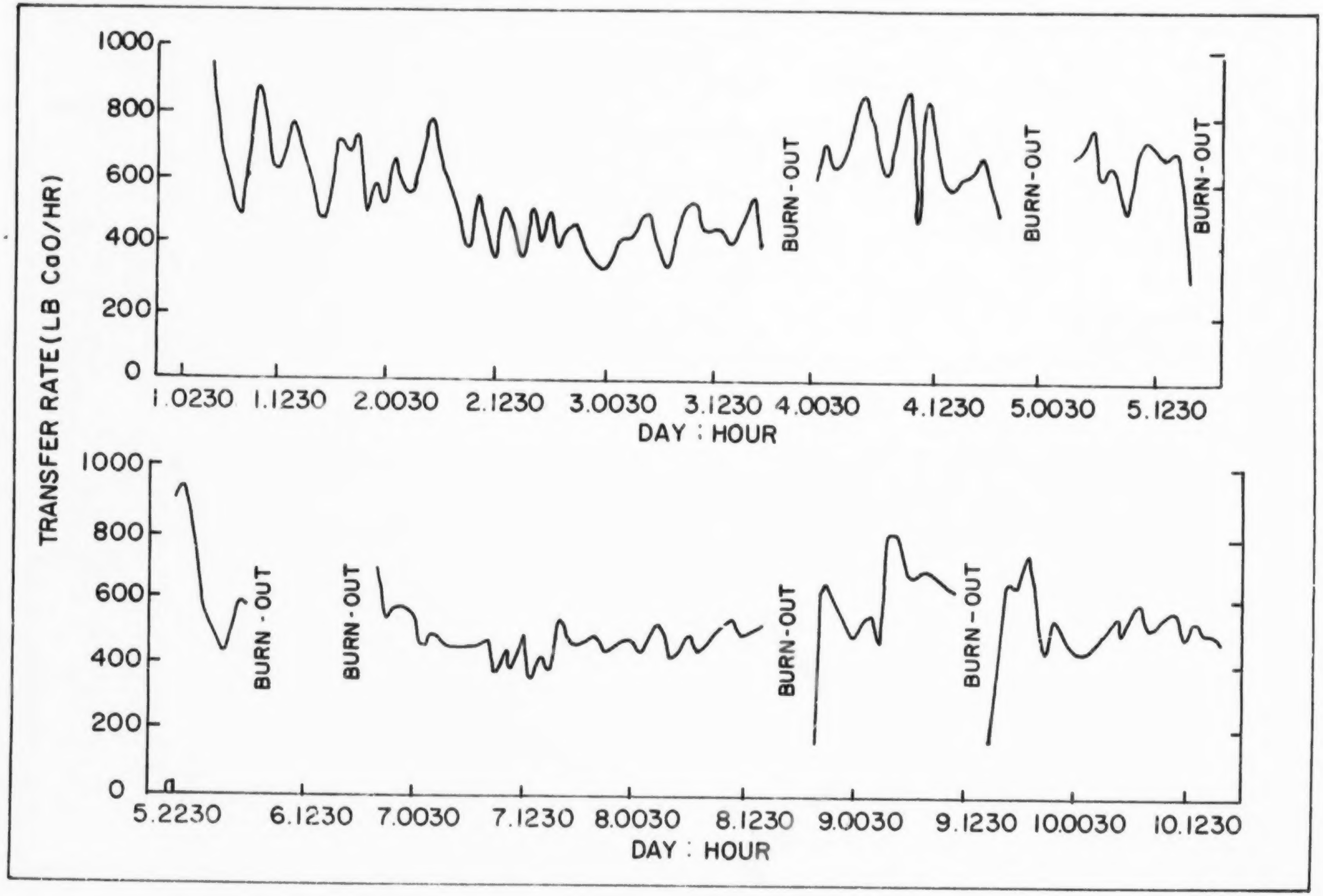


FIG. J-30

CONTINUOUS PILOT PLANT OPERATION LOG (RUN.3.)

SO_2 FROM REGENERATOR

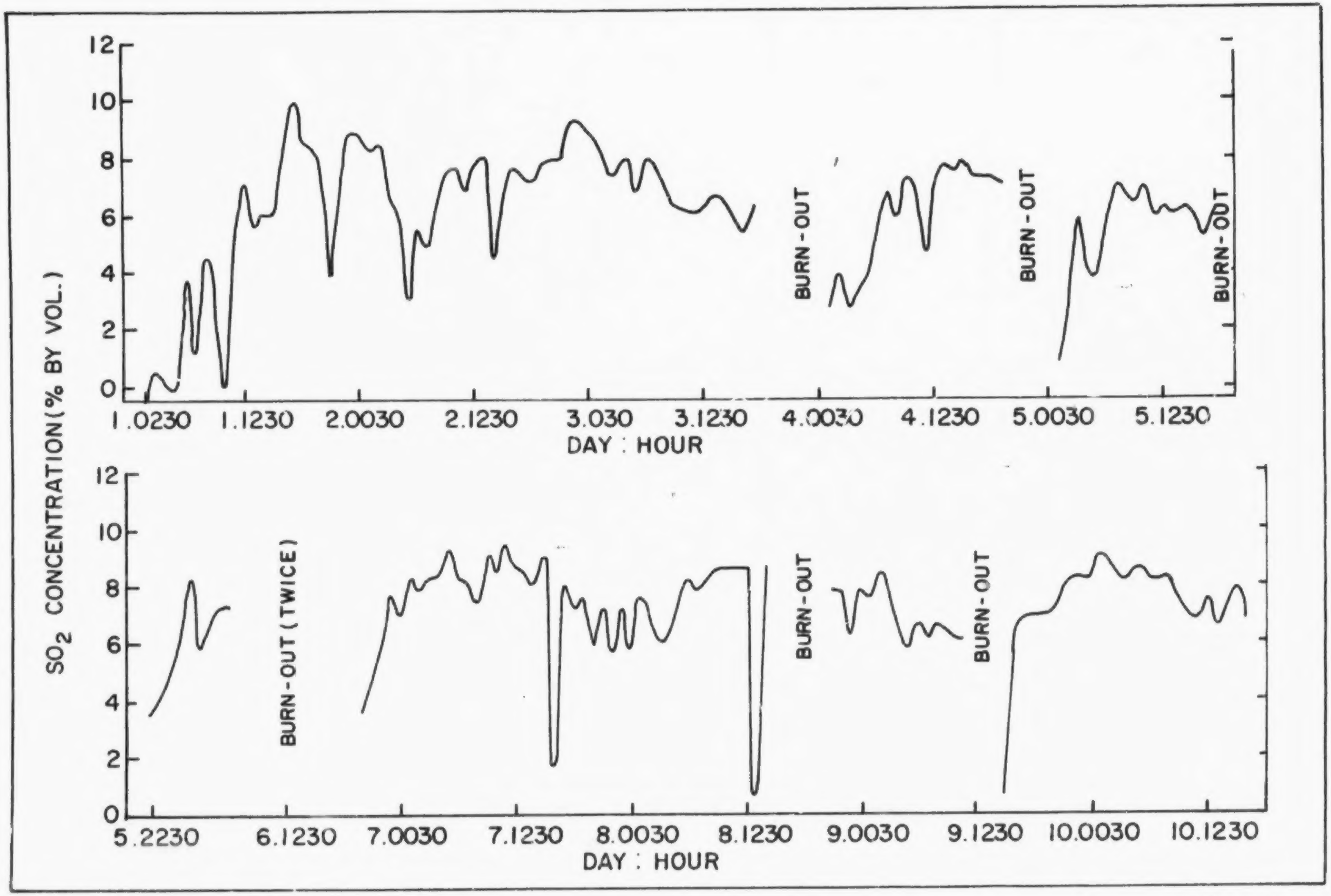


FIG. J-31.

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

CO_2 AND O_2 FROM REGENERATOR

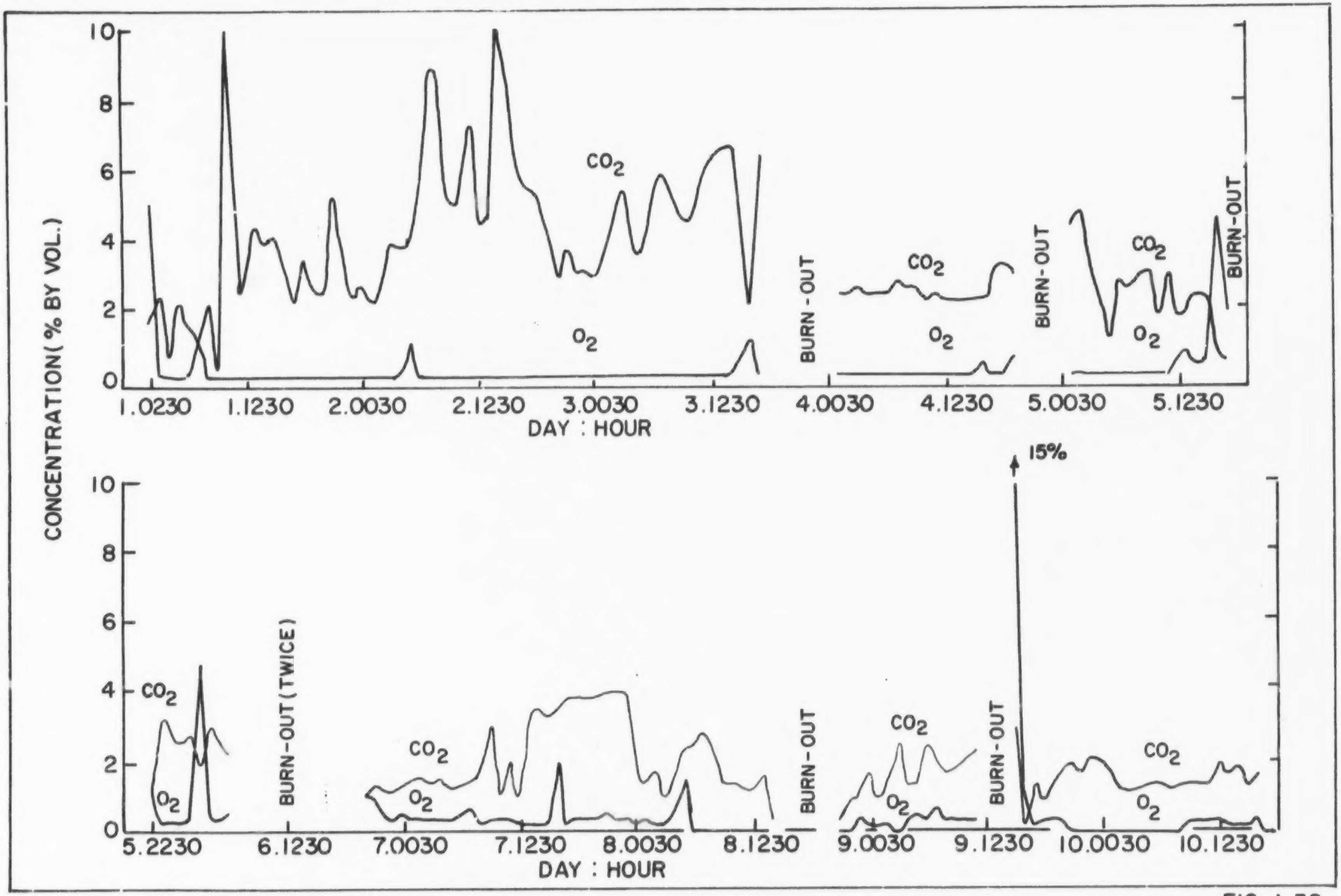


FIG. J-32.

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

SULPHUR FROM REGENERATOR

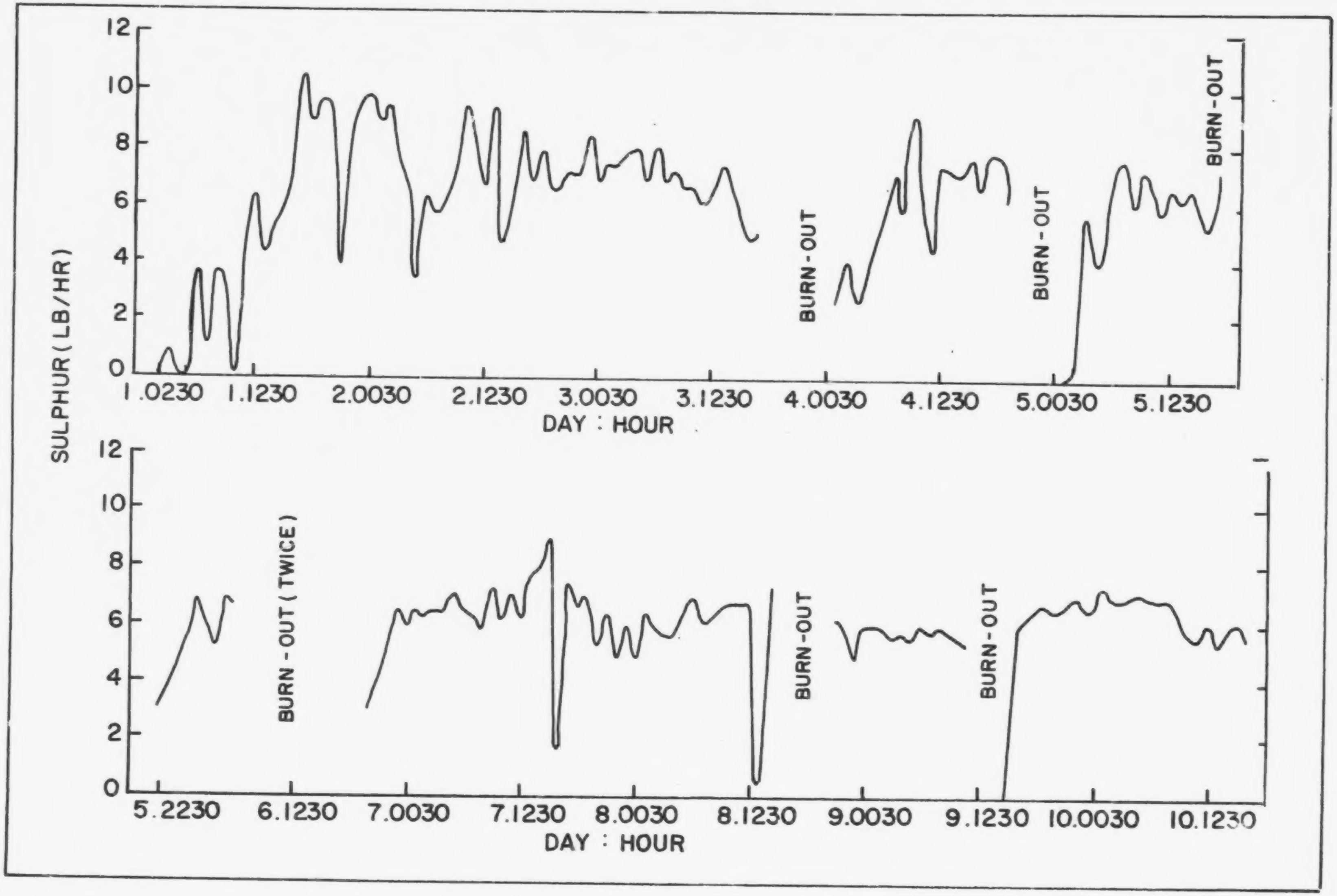


FIG. J-33.

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

SULPHUR FROM REGENERATOR AS % OF SULPHUR FED

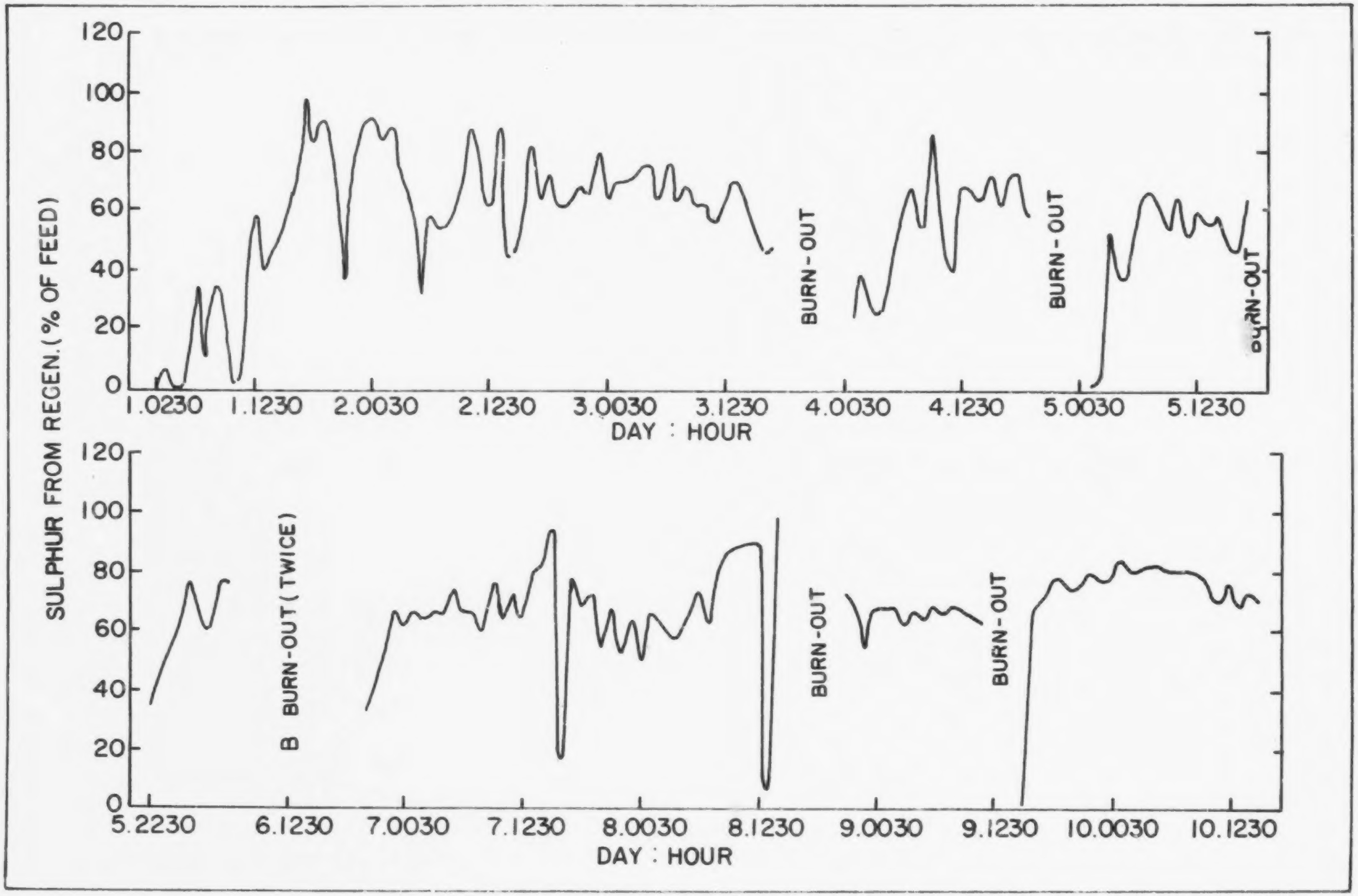


FIG. J-34.

CONTINUOUS PILOT PLANT OPERATION LOG (RUN. 3.)

% CaS CONVERTED TO CaO

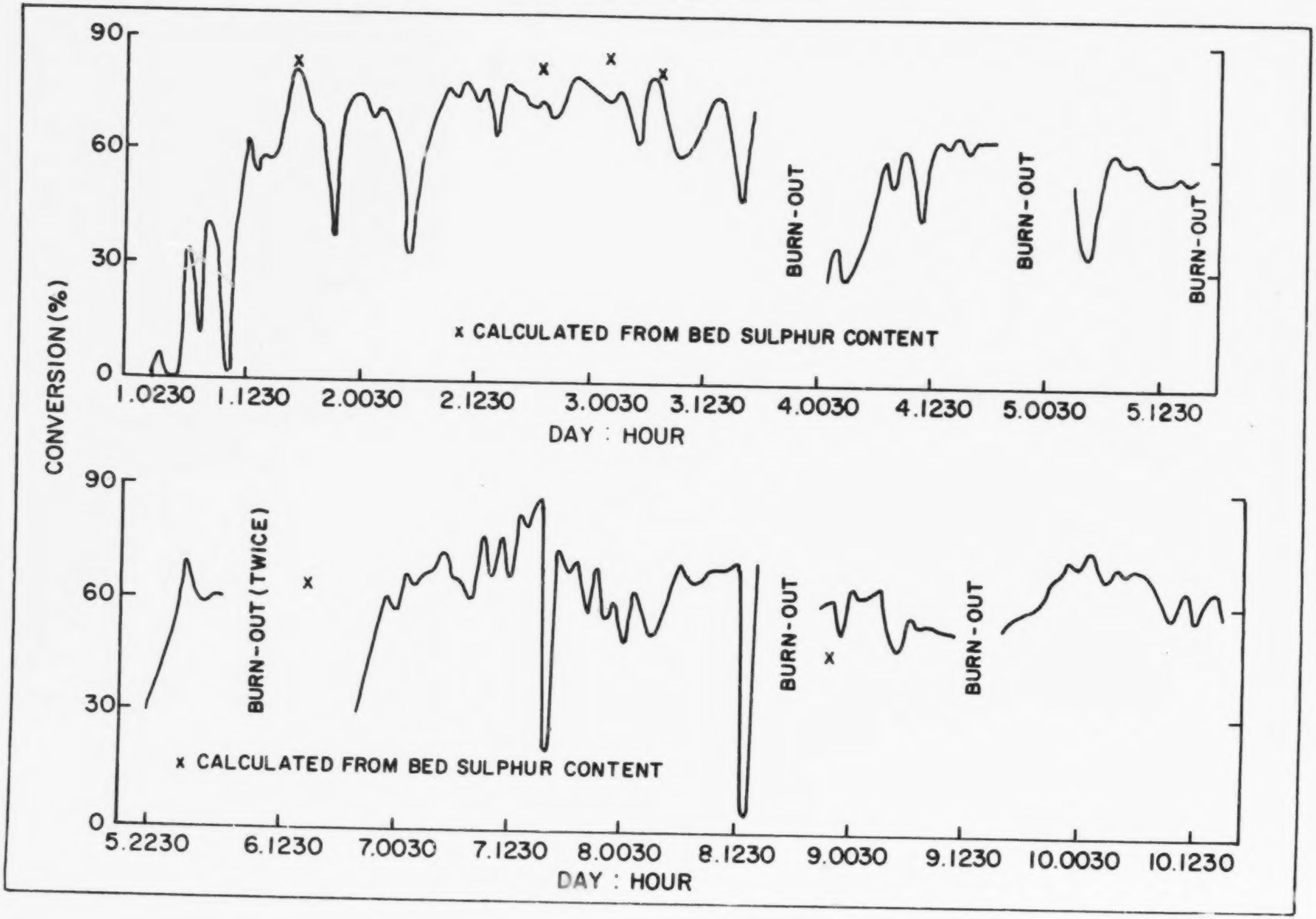
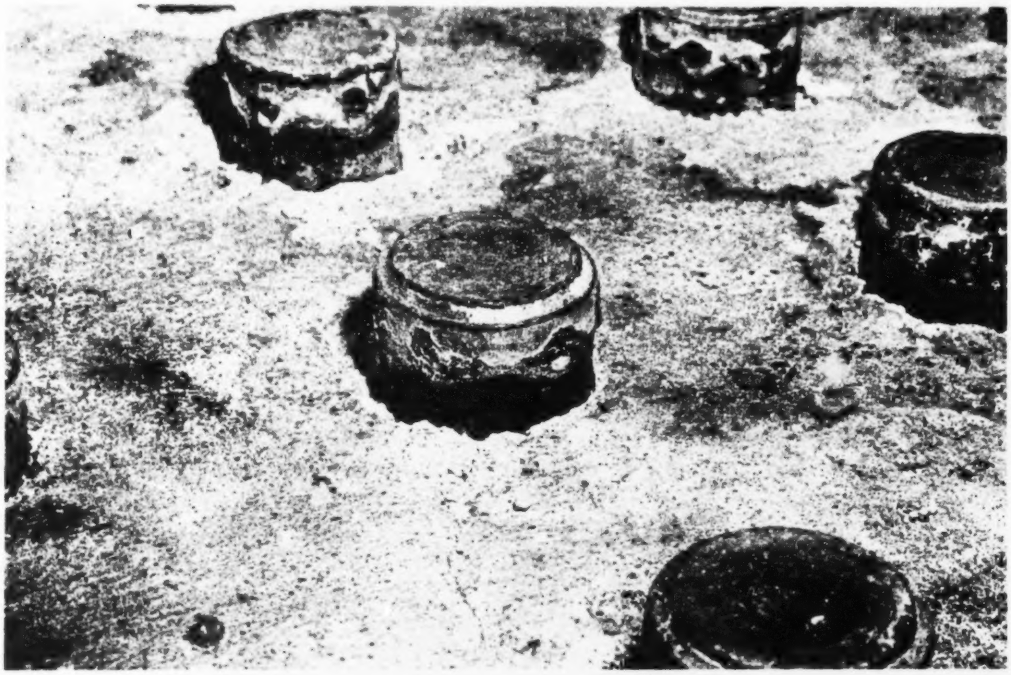


FIG. J-35.

Gasifier Distributor Nozzles After Run 3 Fig J-36



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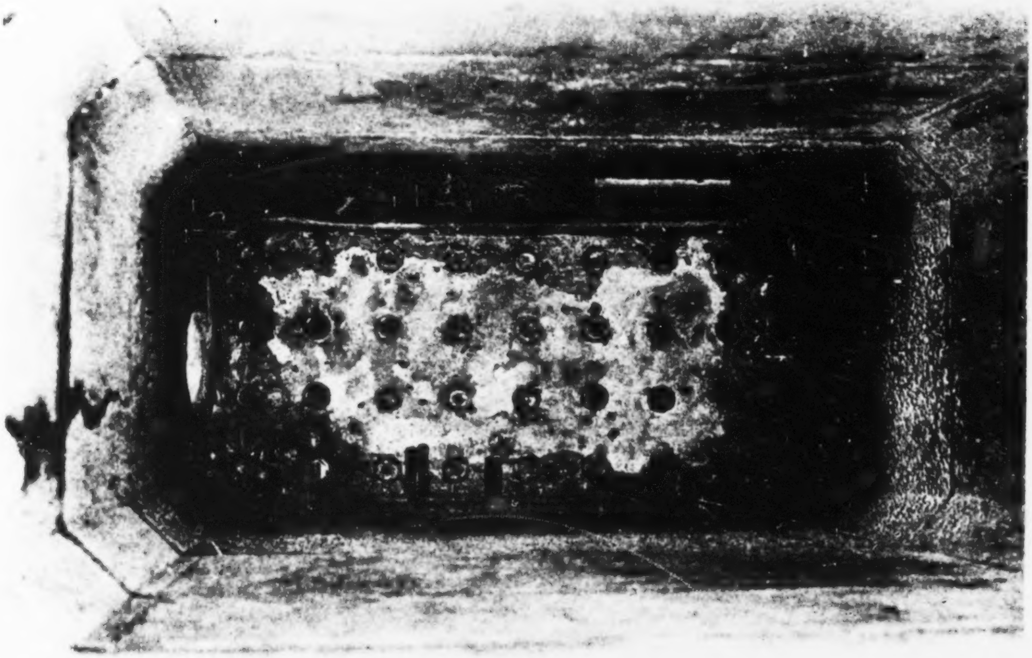
Empty Gasifier Interior After Run 3 Fig J-37



Gasifier Interior, Bed in Place, After Run 5 Fig J-38

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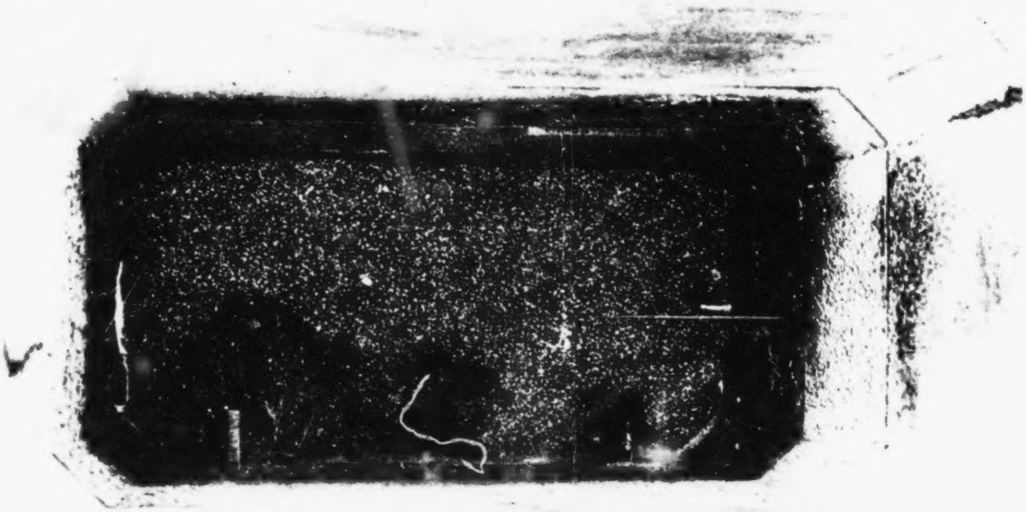
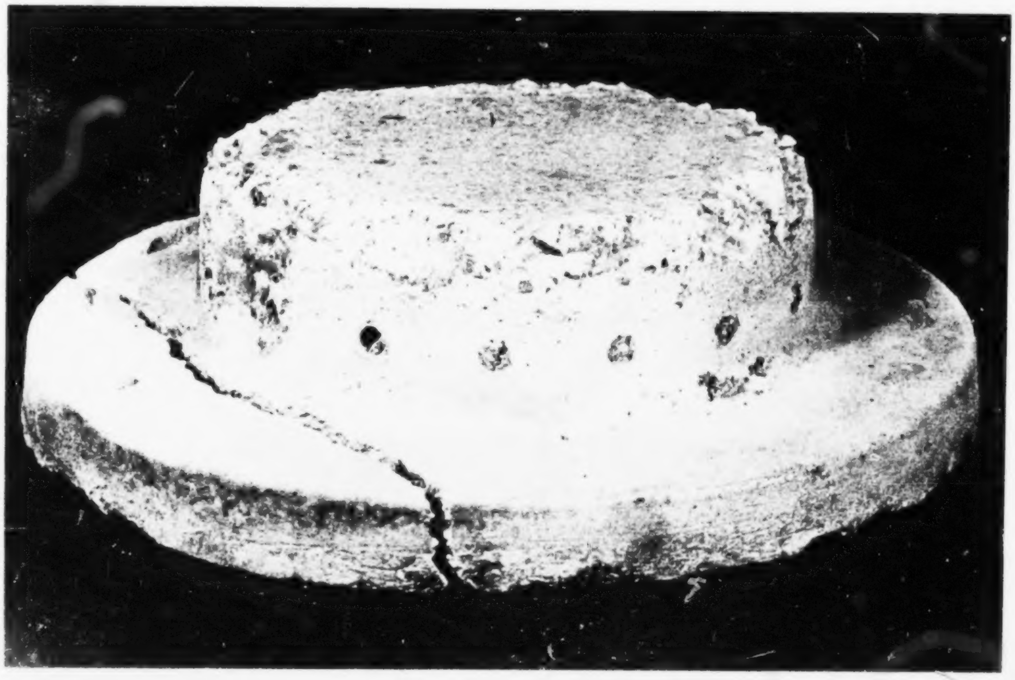
Gasifier Interior, Bed Partly Removed, After Run 5 Fig J-39



Regenerator Distributor After Run 3 Fig J-40



Deposit From Regenerator Bottom After Run 3. Fig J-41

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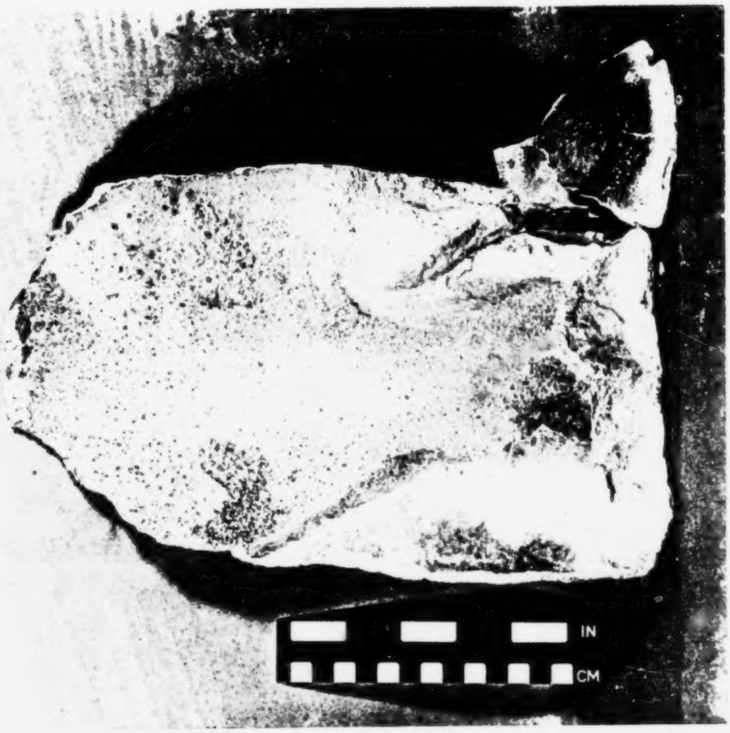
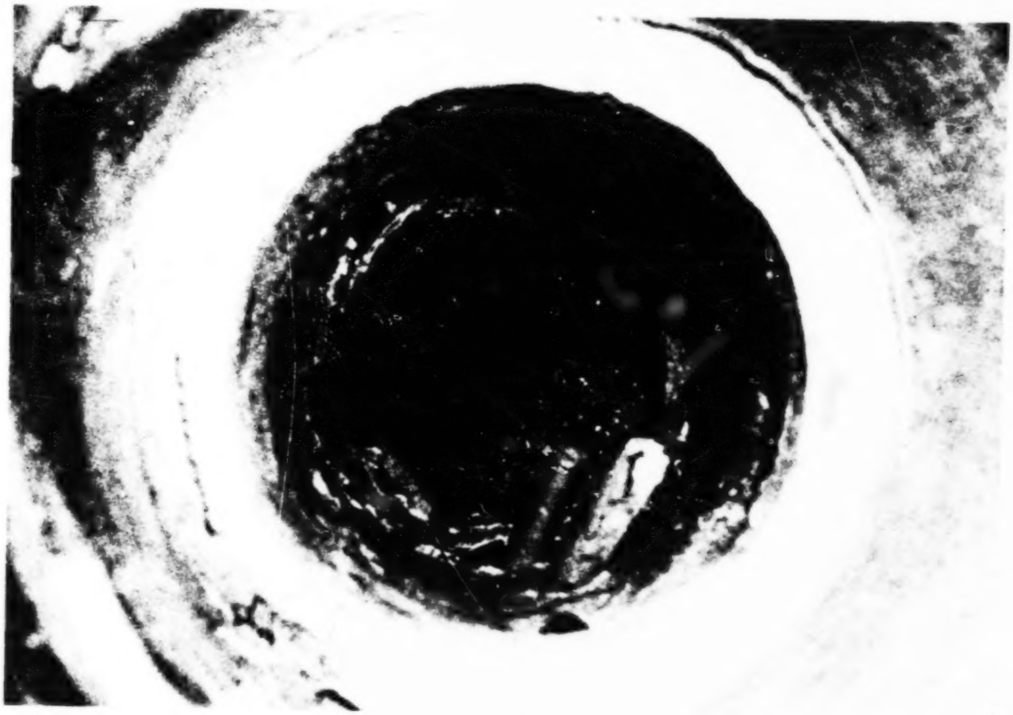
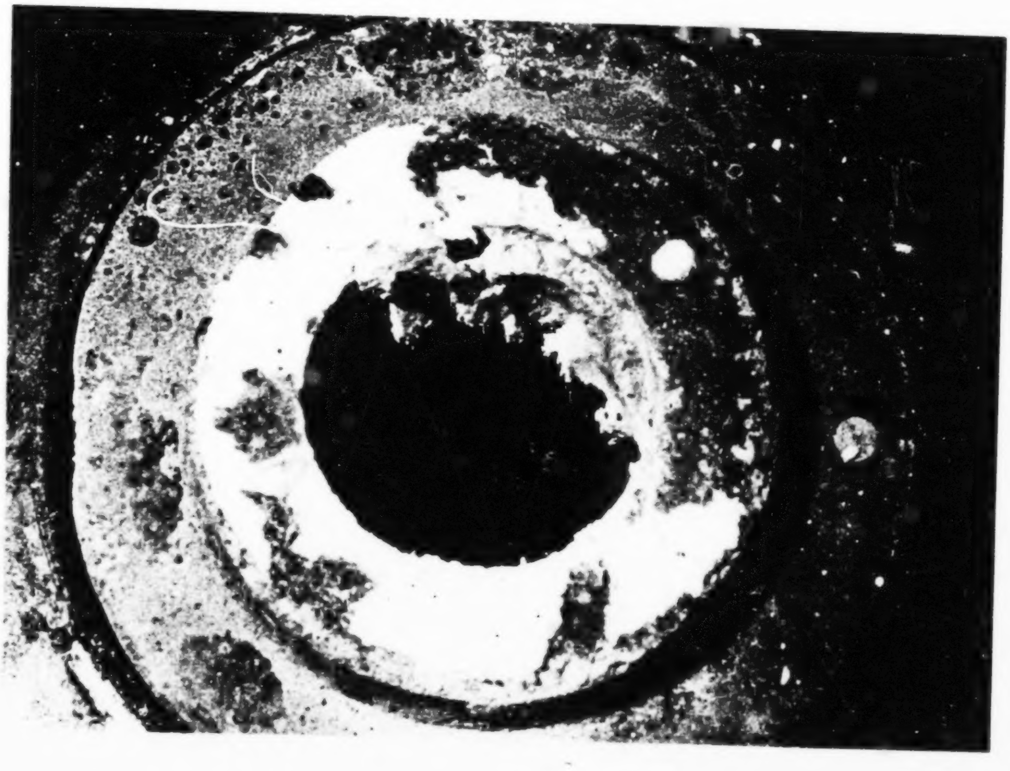


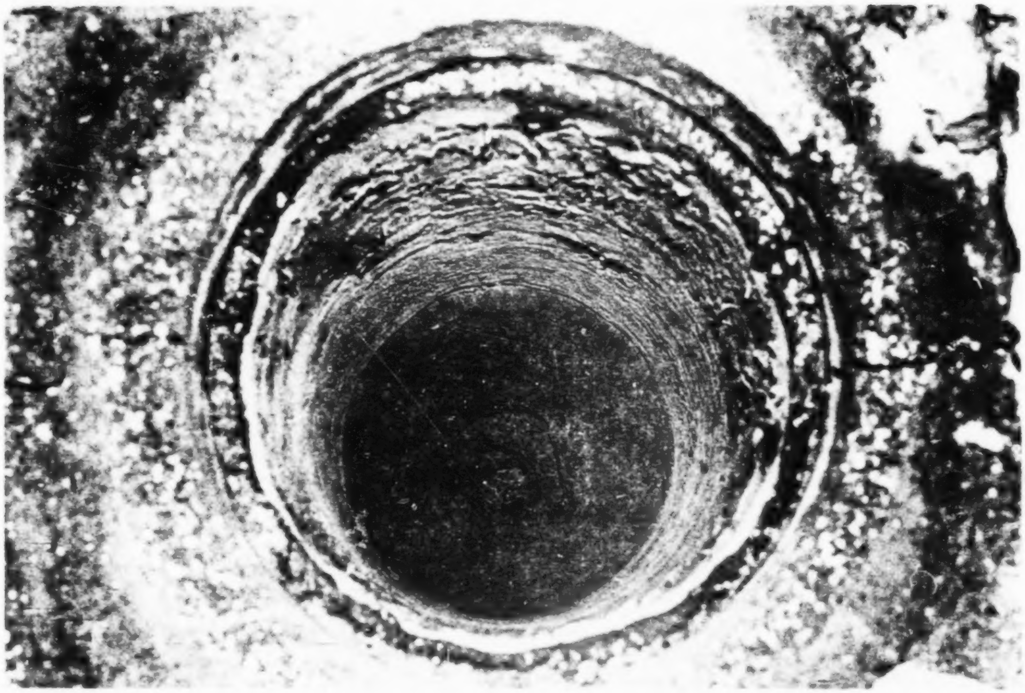
Regenerator Interior From Bottom After Run 3 Fig J-42



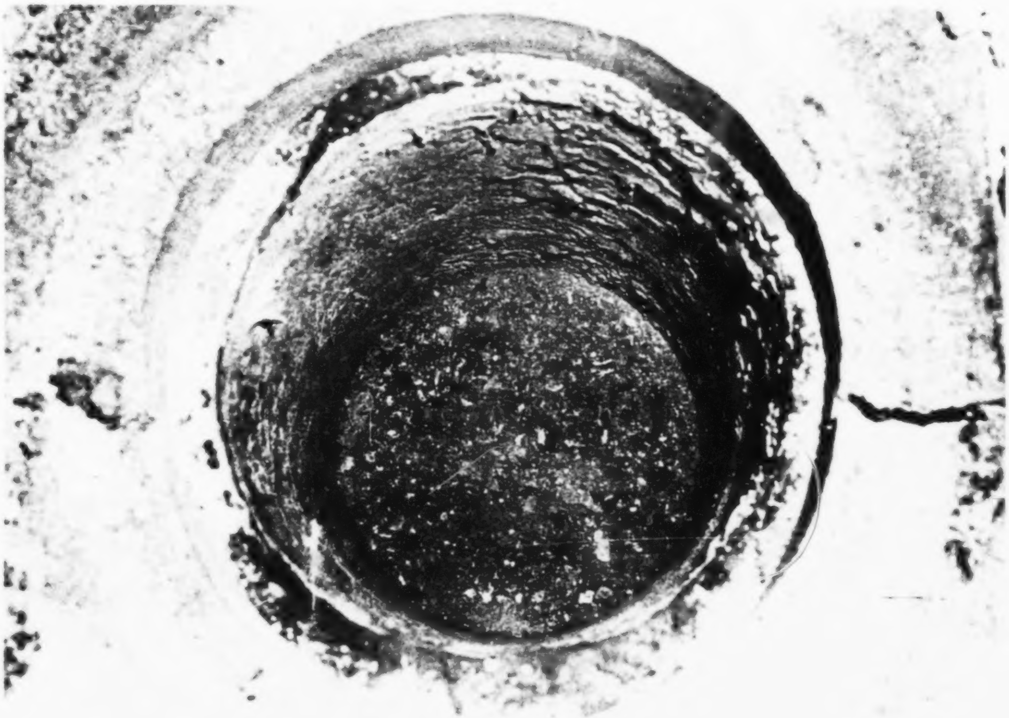
Regenerator Top Gas Outlet After Run 3 Fig J-43



Right Hand Gasifier Cyclone From Top After Run 3 Fig J-44

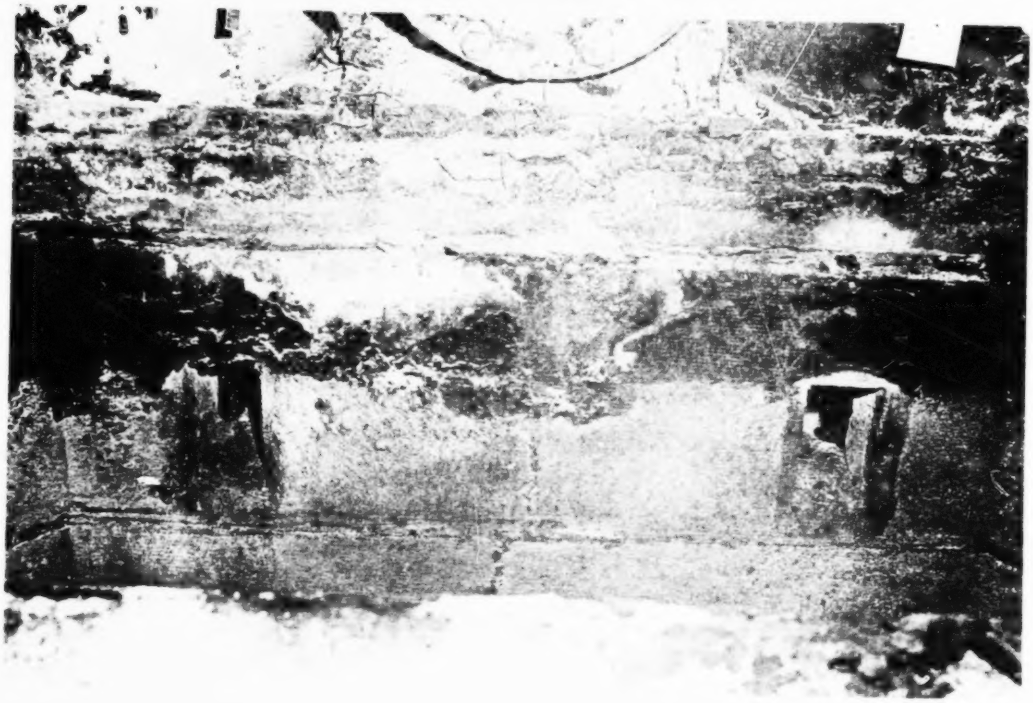


Left Hand Gasifier Cyclone From Top After Run 3 Fig J-45

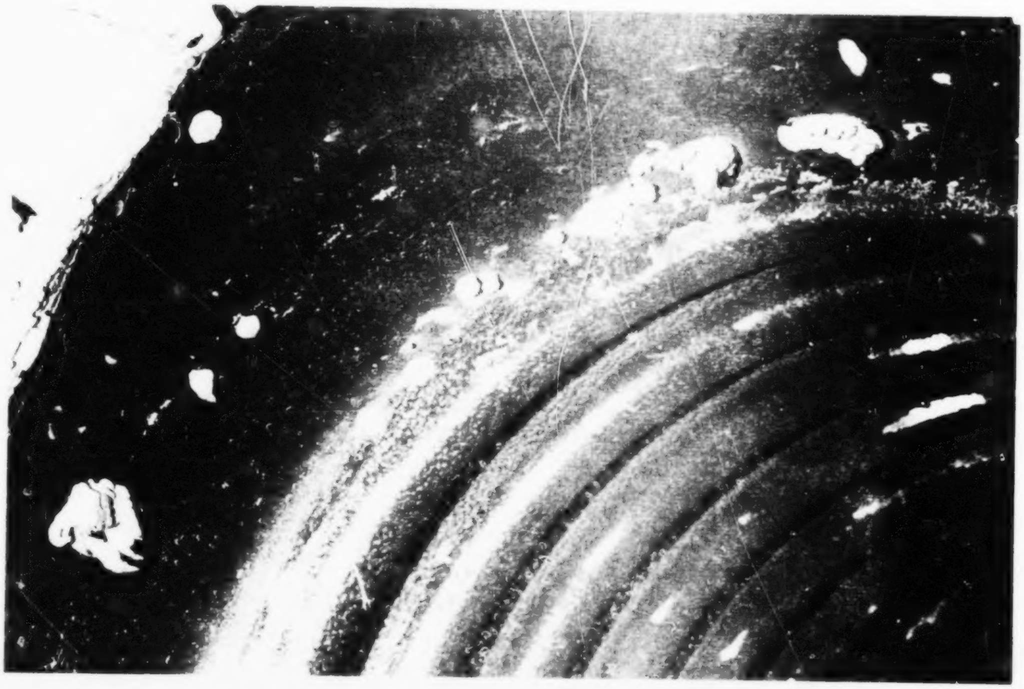


Reproduced from best available copy.

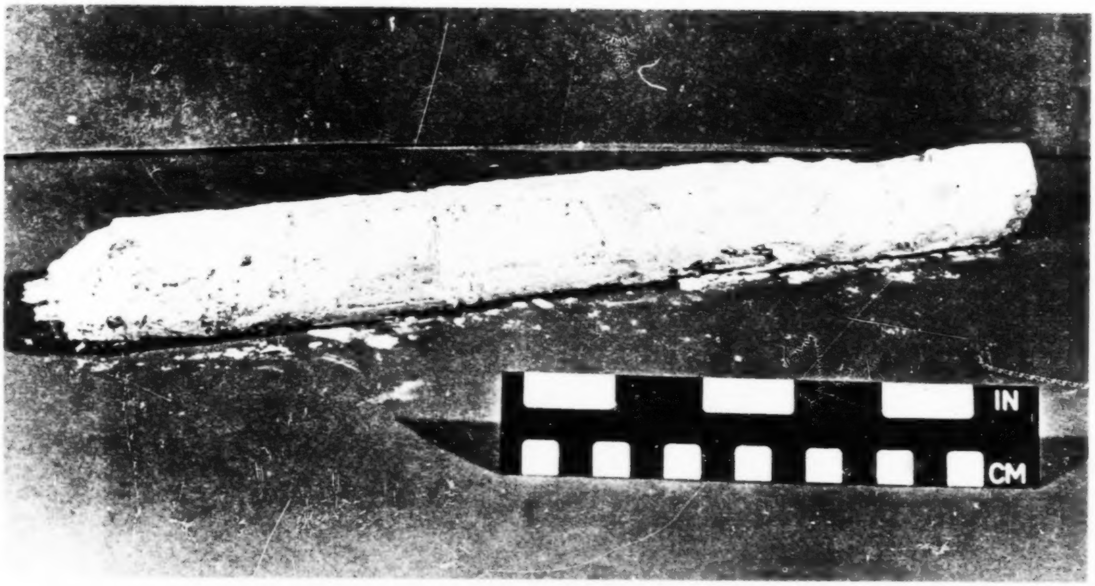
Outlets From Gasifier To Cyclone After Run 3. Fig J-46



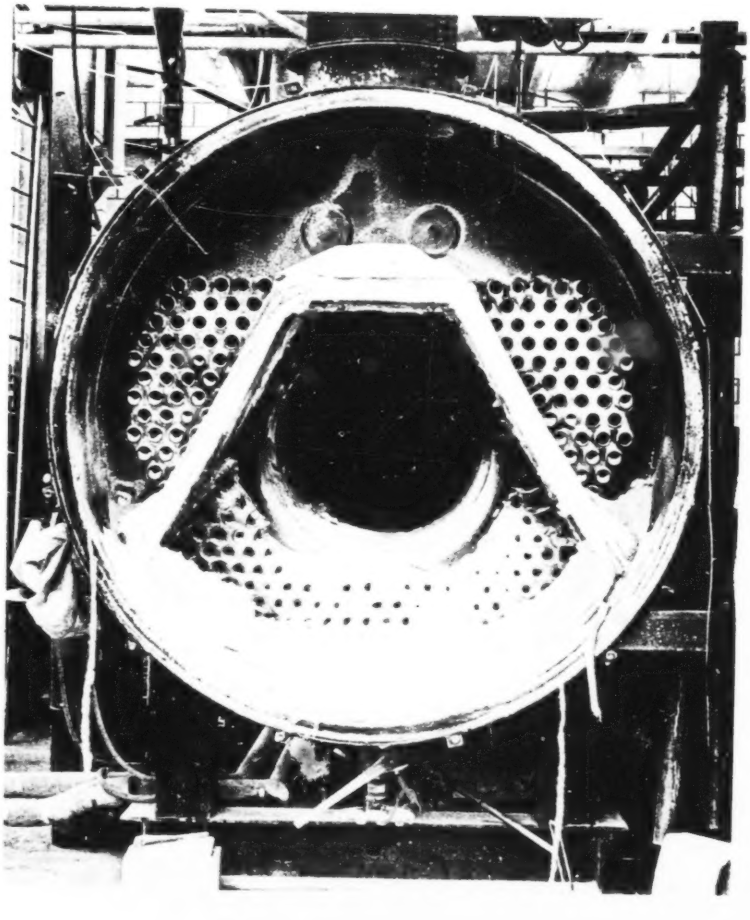
Local Deposits in Boiler Fire Tube After Run 3 Fig J-47



Core of Solids from Boiler Tube After Run 3 Fig J-48



Back End of Boiler After Run 3 Fig J-49



APPENDIX K

BURNING RATE MODEL FOR CARBON ON LIME IN CAFB GASIFIER

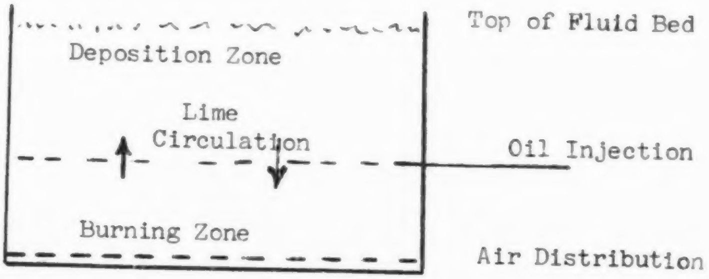
Tests in the CAFB batch unit have revealed an interaction between effects of temperature and air/fuel ratio on the carbon concentration in the lime and on loss in sulphur absorption efficiency.

It has been hypothesized that the concentration of carbon on lime is determined by a dynamic equilibrium between the rate of carbon deposition due to cracking of the fuel and the rate of carbon burning by the air supplied. To test the reasonableness of this hypothesis, a model has been formulated using a simplified geometry and contacting system together with literature values for the rate of carbon oxidation.

Carbon levels computed with this model show order of magnitude correspondence with values actually measured, and show the same type of dependency on temperature and air fuel ratio as observed. The model also predicts an increase in carbon content of lime with increasing superficial air rate at constant temperature and air/fuel ratio. The rapid increase of predicted carbon concentration below 800°C suggests that preheating the gasifier air could improve operations at low air fuel ratio and high gas velocity. The model also suggests that the height of oil injection to the gasifier could be an important variable and warrants further investigation.

A. Physical Model

The physical model assumed is illustrated in Figure K-1.The gasification reactor is assumed to consist of two zones, a carbon burning zone and a carbon deposition zone. The burning zone is below the oil injector, and the deposition zone is above it. It is assumed that oxygen which reaches the deposition zone reacts with hydrocarbon gases and becomes unavailable for coke burning. Each zone behaves as a well mixed reactor for solids but is plug flow in gas. The temperature is constant throughout. Lime circulates between the two zones carrying a layer of coke with it. Coke is burned only in the burning zone. The coke content of the solids is low enough to permit thin layer burning kinetics to apply.



Simplified View of Carbon Burning in Gasifier

Figure K-1

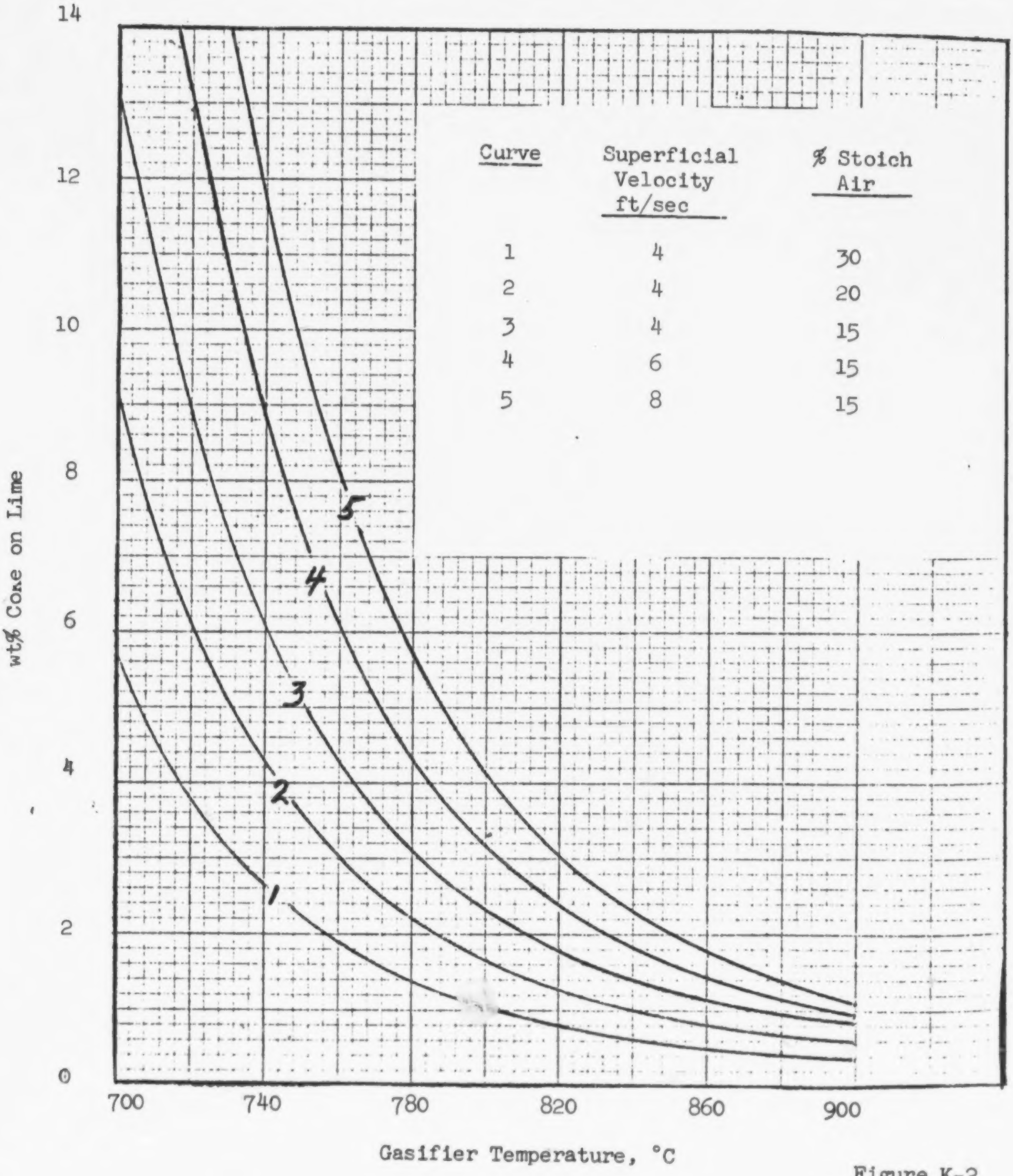
The rate of carbon burning is assumed to follow the first order kinetic equation given by Weisz and Goodwin(8), modified by inclusion of a term for average oxygen concentration. Lime circulation rate is estimated from an empirical equation reported by Kunii and Levenspiel(9). The total coke production rate, equal to the steady state coke burning rate, is taken as the product of oil feed times fraction Conradson Carbon. This assumption is roughly in agreement with fluid coking experience. The average bed coke level is assumed to be the weighted average of the levels in the two zones.

Actual equations employed are described in Section D.

B. Results

Fig.K-2 shows curves of coke concentration vs temperature calculated with the model. The principal feature is the rapid rise of carbon content as temperature falls below about 800°C. The calculations also show a steep rise of carbon content with decreasing air/fuel ratio and with increasing superficial gas velocity. These calculated results are all directionally in agreement with observed behaviour of the CAFB batch reactor results. However the experimental results generally show a somewhat higher increase of carbon with decreasing temperature than calculated.

Computed Coke Concentration on Gasifier Lime



Curve	Superficial Velocity ft/sec	% Stoich Air
1	4	30
2	4	20
3	4	15
4	6	15
5	8	15

Figure K-2

The model contains several parameters which could be adjusted to bring experimental results into better agreement with calculations. For example the boundary between coke burning and deposition zones has been taken as the level of the oil injector. Lowering this boundary would increase the calculated carbon concentrations. The same result would be achieved by a higher coke make than given by Conradson Carbon. It is recommended that additional data on actual coke concentrations be compiled before attempts are made to adjust parameters.

Table III lists values of carbon concentration and recycle rate computed for various conditions. This table shows that differences between predicted coke contents in the two zones are not large in magnitude at the circulation rates estimated. At the higher temperatures, however, the differences do become large relative to the absolute values involved. The calculated circulation rates gives turnover times of the order of a minute. With the circulation rates listed, the average particle residence times in the zones are as follows:

Average Residence Time, Min

Superficial Velocity ft/sec.	Burning Zone	Depositinn Zone
4	0.51	1.34
6	0.28	0.75
8	0.20	0.52

C. Implications

This model assumes a uniform temperature in the carbon burning zone. Obviously the real case is a non uniform temperature distribution due to the cooling effect of the entrance air and the heating effect of the coke combustion. It is likely that the cooling effect reduces the effective volume of the burning zone and increases the carbon concentration. Preheating the gasification air supply should have a beneficial effect on carbon burning and permit operation at lower air/fuel ratios and higher gas velocities.

If the effective boundary between zones is determined by point of oil injection, the location of this point could affect carbon burning. Obviously if the oil were injected too low,the air would contact it before burning much coke. On the other hand, if oil injection is too high there could be insufficient height for sulphur absorption above it. Determination of optimum oil injector location should be made during Phase II studies in the continuous pilot plant.

D. Equations Used in Model

The equations used in the model and their basis are summarized below. Table I lists the symbols and nomenclature used, and Table II is a listing of the computer programme in Basic language. The equations are formulated on the basis of a steady state operating period of the batch reactor.

Coke Input to System

All of the Conradson Carbon of the feed oil is assumed to form coke.

$$\text{Coke input} = F * Cl \qquad (1)$$

Coke Burning Rate

Weisz and Goodwin (8) have reported kinetics for coke burning on catalysts and other solids. It is assumed that the same kinetics apply to burning of coke on lime in the gasifier.

The Weisz and Goodwin equation is

$$\frac{dC}{dt} = -KC \qquad (2)$$

$$\text{where } K = 4 \times 10^7 \ \exp(-37600/R*Tl) \ \text{min}^{-1} \qquad (3)$$

The Weisz and Goodwin work was conducted with air at atmospheric pressure. For the present work it is assumed that the rate is proportional to oxygen concentration, and a term for average oxygen concentration is included.

$$\frac{dC}{dt} = -K * C * Al \qquad (4)$$

Multiplication by the quantity of solids in the burning zone gives an equation for mass rate of coke burning.

$$Sl * \frac{dC}{dt} = -K * C * Al * Sl \qquad (5)$$

Coke Concentration in Burning Zone

Equating the rate of coke input to the rate of coke burning gives an equation for coke concentration in the burning zone.

$$F * Cl = K * C * Al * Sl \qquad (6)$$

$$C = F * Cl/(K * Al * Sl) \qquad (6a)$$

Air Consumption

The coke is assumed to contain hydrogen in addition to carbon. The hydrogen also will consume oxygen. The products of carbon oxidation are CO and CO_2. From experimental results it appears that the composition is kinetically controlled as there is much less CO present than equilibrium would indicate. The grams of oxygen consumed for 100 grams of oil feed is given by:

$$O1 = C1 \ast (H3 \ast 8 + C2 \ast ((C3 \ast 16 + (1-C3) \ast 32)/12)) \qquad (7)$$

The quantity of oxygen supplied per 100 grams of oil is related to the percent of stoichiometric air/fuel ratio by:

$$O2 = 14 \ast S \ast 0.232 \qquad (8)$$

The fraction of oxygen consumed is:

$$O3 = O1/O2 \qquad (9)$$

The log mean average oxygen fraction in the burning zone is:

$$A1 = O3/\text{Log}\ (1/(1-O3)) \qquad (10)$$

Coke Concentration in Deposition Zone

A coke balance into and out of the carbon burning zone provides an equation for the coke concentration in the deposition zone.

Coke transferred in = Coke Burned + Coke transferred out

$$C5 \ast R3 = F \ast C1 + C \ast R3 \qquad (11)$$

$$C5 = (F \ast C1/R3) + C \qquad (11a)$$

Coke Circulation Rate

The vertical circulation of coke across the boundary between zones is related to gas velocity, particle size, and other fluidisation variables. The rate of coke circulation has not been measured for the CAFB case. For this model, the equation of Talmor and Benenati as reported by Kunii and Levenspiel(9) is used. The original equation is:

$$S = 0.785\ (Uo-Umf)\ \text{EXP}\ (-66.3\ dp) \qquad (12)$$

In the nomenclature of this model it becomes:

$$R3 = 60 \ast 0.785 \ast A \ast 30.48\ (V-V1)\ \text{EXP}\ (-.00663D) \qquad (13)$$

Average Coke Concentration

The average coke concentration is determined from the relative heights of the burning and deposition zones by:

$$C6 = (C \ast H1 + C5 \ast (H2-H1))/H2 \qquad (14)$$

TABLE K-I

Symbols and Nomenclature

A	Bed Area sq. cm.
A1	Log mean average oxygen concentration, fraction of that in air.
C	Coke on lime in burning zone, percent.
C1	Conradson carbon of fuel oil, percent.
C2	Carbon in coke, wt. fraction.
C3	CO in carbon oxides from burning zone, mole fraction
C4	Lime charge to bed, grams
C5	Coke on lime in deposition zone, percent
C6	Average coke on lime in bed, percent.
dp	Average lime particle diameter, cm
D	Average lime particle diameter, microns.
F	Oil Feed rate, gm/min.
H1	Height of burning zone, cm
H2	Height of bed, cm
H3	Hydrogen in coke, wt. fraction.
J	Solids circulation rate between zones, gm/cm^2-sec
K	Oxidation rate constant, min^{-1} (K=R2)
O1	Oxygen consumption rate, gm/100 gm oil feed.
O2	Oxygen feed rate, gm/100 gm oil feed.
O3	Oxygen consumption, fraction of O2 fed
R	Gas Constant, Cal/g Mole - Deg kelvin.
R1	Gas density, lb/cu-ft.
R2	Oxidation rate constant, min^{-1} (R2=K).
R3	Solids circulation rate, gm/min.
S	Air fuel ratio, percent of stoichiometric.
S1	Solids in burning zone, g.
t	Time, minutes.
T	Temperature, Deg C.
T1	Temperature, Deg K.
T2	Temperature, Deg R.
Uo	Superficial Gas velocity, cm/sec.
Umf	Minimum fluidisation velocity, cm/sec.
V	Superficial gas velocity, ft/sec
V1	Minimum fluidisation velocity, ft/sec.

TABLE K-II

Computer Programme for Carbon Calculation

```
100 DATA 4, 700, 15, 12.7, 46
110 DATA 8240, 248, 13., .9796, .0204
120 DATA .5 , 800, 1.5
130 READ V, T, S, H1, H2, C4, A, C1, C2, H3, C3, D, V1
140 REM SUPERFICIAL VELOCITY, TEMP C, FRACTION STOIC AIR,
150 REM BURN ZONE HEIGHT, CM.
160 REM LIME CHARGE GM, AREA SQ CM, PCT CON CARBON, FR C IN COKE
170 REMFR H IN COKE
180 REM CO FRACTION, PARTICLE DIA MICRONS, MIN FLUID VELOCITY
190 PRINT"CALCULATION OF CARBON ON STONE"
200 PRINT"                FOR"
210 PRINT USING 610,H1,H2
220 PRINT USING 620, C4, A
230 PRINT USING 630,C1,C2
240 PRINT USING 640,H3,C3
250 PRINT USING 650, D, V1
260 PRINT
270 FOR V=4 TO 8 STEP 2
280 FOR S=15 TO 30 STEP 5
290 PRINT USING 300,V,S
300:  VELOCITY=###.# FT/SEC.   AIR/FUEL RATIO=####.# PERCENT STOICH.
310 PRINT
320 PRINT USING 580
330 PRINT USING 590
340
350 FOR T = 700 TO 900 STEP 20
360 LET T1=T+273.16
370 LET T2=1.8*T1
380 LET O1=C1*(H3*8+C2*((C3*16+(1-C3)*32)/12))
390 LET O2=14*S*.232
400 LET O3=O1/O2
410 LET A1=O3/LOG(1/(1-O3))
420 LET S1 = (H1/H2)*C4
430 LET R1=29*520/(380*T2)
440 LET A2 = V*A*.001073
450 LET A3=A2*R1*60*454
460 LET F=A3/(.14*S)
470 LET R2=4E7*EXP(-37600/(1.987*T1))
480 LET C = F*C1/(R2 * S1 * A1)
490 LET R3 = 60*.785*A*30.48*(V-V1)*EXP(-.00663*D)
500 LET C5 = (F*C1/R3) + C
510 LET C6 = (C*H1 + C5*(H2-H1))/H2
520 PRINT USING 600, T, C, C5, C6, R3
530 NEXT T
540 PRINT
550 PRINT
560 NEXT S
570 NEXT V
580:   TEMP   PCT. CARBON IN LIME  RECYCLE
590:    C    BURN  DEPOSIT  TOTAL  GM/MIN
600:   ###.  ##.##  ##.##   ##.##  #####.
610:   FEED HEIGHT   = ####.# CM        BED HEIGHT  = ####.# CM
620:   CHARGE(LIME) = #####.# GRAMS,   BED AREA    = ####. SO CM
630:   CON.CARBON    =####.# PERCENT, FUEL CARBON = ###.#### FRACT.
640:   FUEL HYDROGEN =   #.#### FR.,  CO FRACTION = ###.###
650:  PARTICLE DIA = ####. MICRONS, MIN FLUID VEL = ##.## FT/SEC
660 END
```

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TABLE K-111

Summary of Computed Results

CONDITIONS

FEED HEIGHT =	12.7	BED HEIGHT =	46.0 CM
CHARGE (LIME) =	8200.0 GRAMS	BED AREA =	248. SQ.CM
CON CARBON =	10.0 PERCENT	COKE CARBON =	.9796 FR.
COKE HYDROGEN =	.0204 FR.,	CO FRACTION =	.500
PARTICLE DIA =	800. MICRONS,	MIN FLUID VEL =	1.50 FT/SEC

GAS VELOCITY FT/SEC	STOICH AIR PCT	TEMP C	PERCENT COKE ON LIME B ZONE	D ZONE	TOTAL	RECYCLE GM/MIN
4	15	700.	12.63	13.34	13.14	4425
"	"	800.	1.87	2.51	2.34	"
"	"	900.	.38	.97	.81	"
"	20	700.	8.74	9.27	9.12	"
"	"	800.	1.29	1.77	1.64	"
"	"	900.	.26	.70	.58	"
"	25	700.	6.69	7.11	7.00	"
"	"	800.	.99	1.38	1.27	"
"	"	900.	.20	.55	.46	"
"	30	700.	5.42	5.78	5.68	"
"	"	800.	.80	1.12	1.04	"
"	"	900.	.16	.46	.38	"
6	15	700.	18.95	19.54	19.37	7965
"	"	800.	2.81	3.34	3.19	"
"	"	900.	.57	1.06	.92	"
"	20	700.	13.10	13.55	13.42	"
"	"	800.	1.94	2.34	2.23	"
"	"	900.	.39	.76	.66	"
"	25	700.	10.04	10.39	10.29	"
"	"	800.	1.49	1.81	1.72	"
"	"	900.	.30	.60	.51	"
"	30	700.	8.14	8.43	8.35	"
"	"	800.	1.21	1.47	1.40	"
"	"	900.	.25	.49	.42	"
8	15	700.	25.26	25.81	25.66	11505
"	"	800.	3.74	4.24	4.10	"
"	"	900.	.76	1.21	1.09	"
"	20	700.	17.47	17.88	17.77	"
"	"	800.	2.59	2.96	2.86	"
"	"	900.	.53	.86	.77	"
"	25	700.	13.38	13.71	13.62	"
"	"	800.	1.98	2.28	2.20	"
"	"	900.	.40	.67	.60	"
"	30	700.	10.85	11.12	11.05	"
"	"	800.	1.61	1.85	1.79	"
"	"	900.	.33	..55	.49	"